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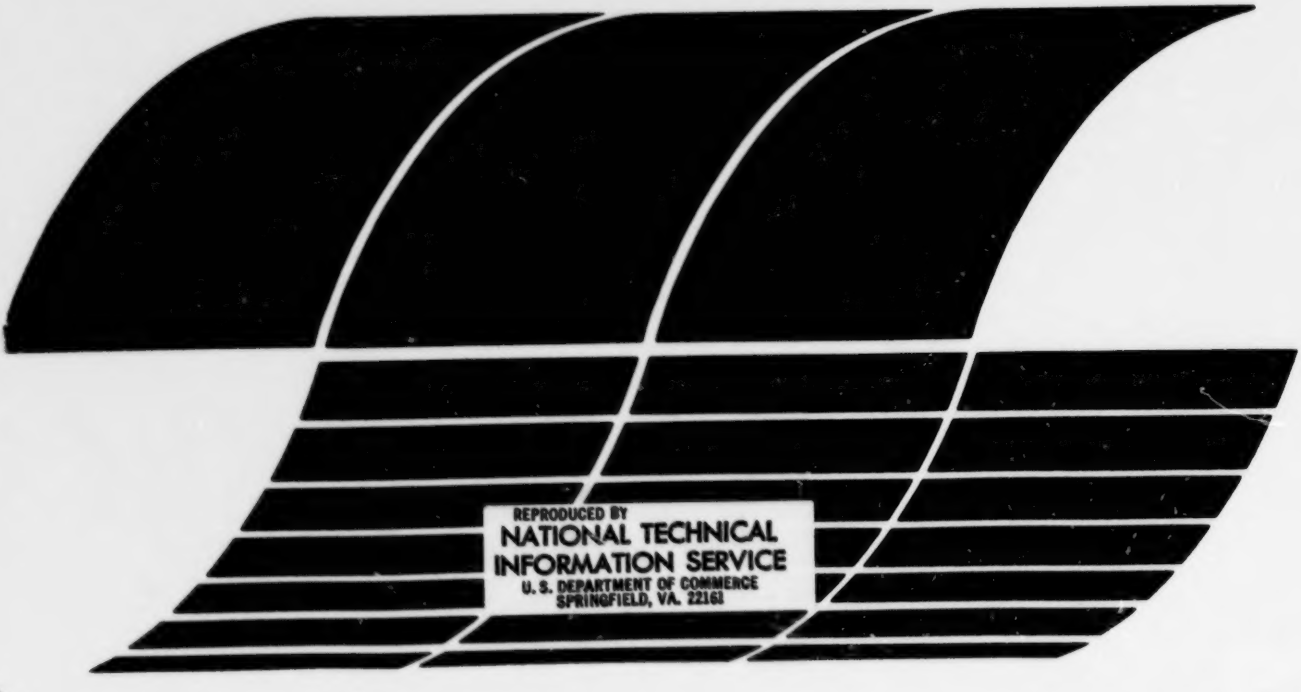

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a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Desulfurization	Pollution Control	13B	07D
Coal Gasification	Nitrogen Oxides	Stationary Sources	13H	07B
Gas Purification	Dust	Environmental Impact	07A	11G
Fuels	Chemical Analysis	Fuel Gas	21D	
Toxicity		Particulate	06T	
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ABSTRACT

United Technologies Research Center continues to demonstrate the potential environmental and economic benefits of integrated coal gasification/gas cleanup/combined gas and steam cycle power plants. Several technical problem areas requiring investigation or further definition were left over from UTRC's work on EPA Contract No. 68-02-1099, reported in EPA-600/2-76-153 (June 1976). Refinements in plant operational characteristics lower heat rates and reduce emissions from previous values. An expanded study of plant environmental intrusions includes a look at potentially hazardous trace elements. Comparisons made of integrated plants using air-blown and oxygen-blown gasifiers favor air-blown operation. Careful theoretical design of plants with low temperature sulfur cleanup reduces to marginal levels the performance and cost advantages of plants with high temperature cleanup. If gasifier steam feed rates are kept low in all but fixed bed types, choice of gasifier among other major generic types is not critical to achieving attractive systems using low temperature cleanup. Excessive thermal NO_x emissions may be avoided by departing from conventional combustor designs. Fuel NO_x and particulates remain as unsolved problems with use of high temperature cleanup. Sulfur removal to very low levels is possible with integrated systems, but cost rises rapidly as it becomes necessary to remove most of the COS as well as the H_2S.

Fuel Gas Environmental Impact

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CONVERSION FACTORS

ft x 0.3048 = m
ft^3 x .02832 = m^3
gal x .003785 = m^3
GPM x 4.403 = m^3/s
atm x 101,325 = N/m^2 (Pa)
lb_f/in^2 x 6895 = N/m^2 (Pa)
lb_f x 4.448 = N
lb_m x .4536 = kg
ton x 907.2 = kg
hp x 746 = W
Btu x 1056 = J
*Btu/ft^3 x 37,288 = J/m^3
Btu/lb mol x 2328 = J/kg mol
$lb_m/10^6$ Btu x .4295 = $kg/10^9$ J
F subtract 32 x 5/9 = C

---

* Differences between various "standard" conditions (e.g., SCF vs. normal m^3) may require slight modifications to this conversion factor.

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Technical Project Officers for the EPA were Mr. W. J. Rhodes until July 1977 and, subsequently, Dr. T. W. Petrie. Valuable guidance and comments received from these gentlemen are gratefully acknowledged.

Process information was made available from current studies sponsored by the Electric Power Research Institute (EPRI). The cooperation and comments of Dr. M. J. Gluckman of EPRI are also gratefully acknowledged.

SECTION 1

SUMMARY

INTRODUCTION

The electric utility industry is the largest single consumer of basic energy in the United States. As such, it is under massive pressure to utilize indigenous fuels, especially coal, the most plentiful fossil fuel. However, the industry realizes that environmental protection is a primary concern and, thus, there is a great interest in methods of generating electricity with minimum environmental intrusion. One of the first studies to identify advanced methods of generating electric power at minimum emissions and cost was reported on in December 1970 (Reference 1-1) by the United Technologies Research Center (UTRC) for the National Air Pollution Control Administration, an EPA predecessor organization. In this study, the combined gas turbine-steam turbine power cycle, or simply combined-cycle system, used in conjunction with coal gasification and fuel gas cleanup (Figure 1-1) was identified as the most attractive of the various advanced power systems in terms of emissions and cost of power. Subsequent studies by a wide variety of organizations (e. g., References 1-2, 1-3, and 1-4) have reached the same conclusion.

Since 1973, UTRC has been carrying out further studies for the EPA Fuel Process Branch, Industrial Environmental Research Laboratory/Research Triangle Park with the objective of providing more detailed definitions of the technological status, the potential environmental intrustion, the performance, and economics of coal gasification with both low- and high-temperature sulfur cleanup processes integrated with combined-cycle power systems. The model integrated power plants are nominal 1000-MW, base-load plants.

In this and previous studies (References 1-5 and 1-6) fixed-bed, fluid-bed, entrained-flow and molten salt gasifiers have been considered as have various types of physical and chemical sorbent cleanup systems. Power systems technology has been both first-generation (~1980-82) systems operating at 2200 F* turbine inlet and second-generation (post 1985) systems

---

*EPA's policy is to use metric units. For the convenience of the reader making comparisons to results of previous studies, non-metric units are used herein. A list of conversion factors to metric units is given on page xiv.

TYPICAL PRESENT–DAY WASTE–HEAT COMBINED GAS AND STEAM TURBINE SYSTEM

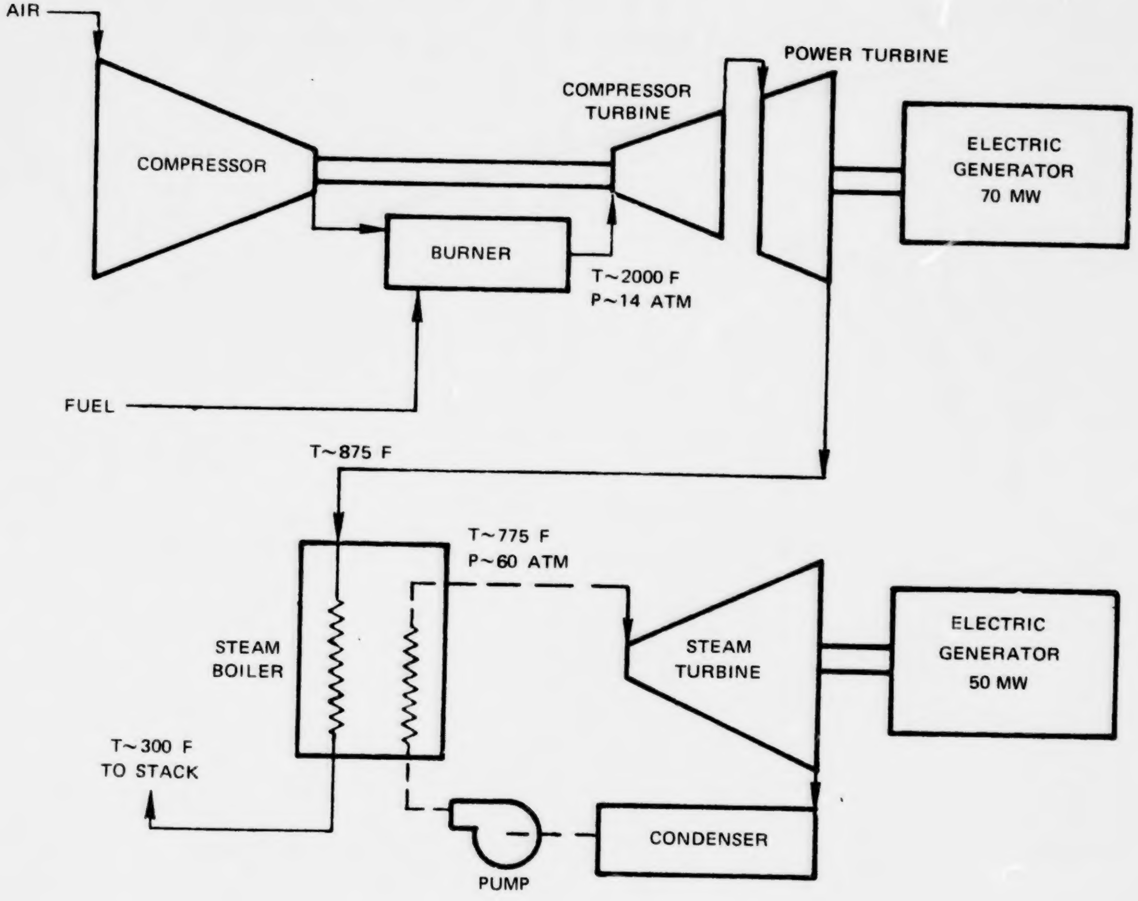

AIR
COMPRESSOR
BURNER
FUEL
COMPRESSOR TURBINE
T~2000 F
P~14 ATM
POWER TURBINE
ELECTRIC GENERATOR 70 MW
T~875 F
STEAM BOILER
T~775 F
P~60 ATM
STEAM TURBINE
ELECTRIC GENERATOR 50 MW
T~300 F TO STACK
PUMP
CONDENSER

FIG.1–1

operating at 2600 F. Emphasis has been placed on the latter since it appears that it will be the last half of the 1980 decade before gasification processes become available for utility use.

In the performance of these studies, UTRC has had the expert assistance of Foster Wheeler Energy Corporation, Fluor Engineers and Constructors, Inc., Hittman Associates, Inc. and the Allied Chemical Company. In addition, the Electric Power Research Institute (EPRI) has provided considerable data on several gasification processes which were the subject of current EPRI-sponsored research at Fluor Engineers and Constructors. The Conoco Coal Development Co. has provided data relating to their half-calcined dolomite desulfurization process.

The major conclusion of these studies is that the integrated coal gasification/sulfur cleanup/combined-cycle power system offers the potential of lower environmental intrusion and lower cost electricity than conventional power plants with flue gas desulfurization. During the course of the studies, the gasifier operating conditions were updated and modified where possible to reduce utility consumption and thereby increase the gasifier efficiency. This is reflected throughout the entire power plant by lower capital and fuel costs and lower emissions. Thus, each subsequent study phase has indicated an increased attractiveness for the integrated power plants.

For the systems studied, sulfur removal costs showed little increase as emissions were reduced to one-third or less of the EPA standard for coal-fired power plants. While fuel gas cleanup equipment for medium-Btu gas is smaller and less costly than for low-Btu gas, an integrated plant with an oxygen-blown gasifier showed no advantage over its air-blown counterpart; in fact, for the gasifier type considered, it had a higher heat rate, higher power costs, and potentially higher NO_x emissions.

During the course of the overall program, continual refinements were made in the integration process which were more advantageous to the systems using low-temperature cleanup. Thus, early estimates for the performance advantages of the integrated power plant with high-temperature sulfur removal have been decreased to marginal levels. Since the performance and cost of the high-temperature cleanup system are based upon extrapolations of bench-scale tests, whereas those of the low-temperature system are based upon adaptations of commercial-scale equipment, it would not be unfair to presume that the proven low-temperature systems should offer an attractive choice.

POWER PLANT COMPONENTS

The power plant consists of three basic elements: gasification, cleanup and power generation. One of the most important factors in the gasified coal/combined-cycle power system is the integration of the components to achieve proper utilization of waste heat both in the steam cycle and in the thermal regeneration of the fuel gas streams.

Gasification

The major generic types of gasifiers have been examined including fixed, entrained, and fluidized bed designs and the molten salt gasifier which combines gasification and sulfur removal in a sodium carbonate melt. Table 1-1 lists the gasifiers considered. Emphasis has been placed on those gasifier operating characteristics, especially utility consumption, which are a key factor in performance of the cleanup system and the combined-cycle generating plant. For example, one change which has a widespread effect, particularly on the air-blown, entrained-bed gasifier (Figure 1-2) is the reduction in steam feed rate.

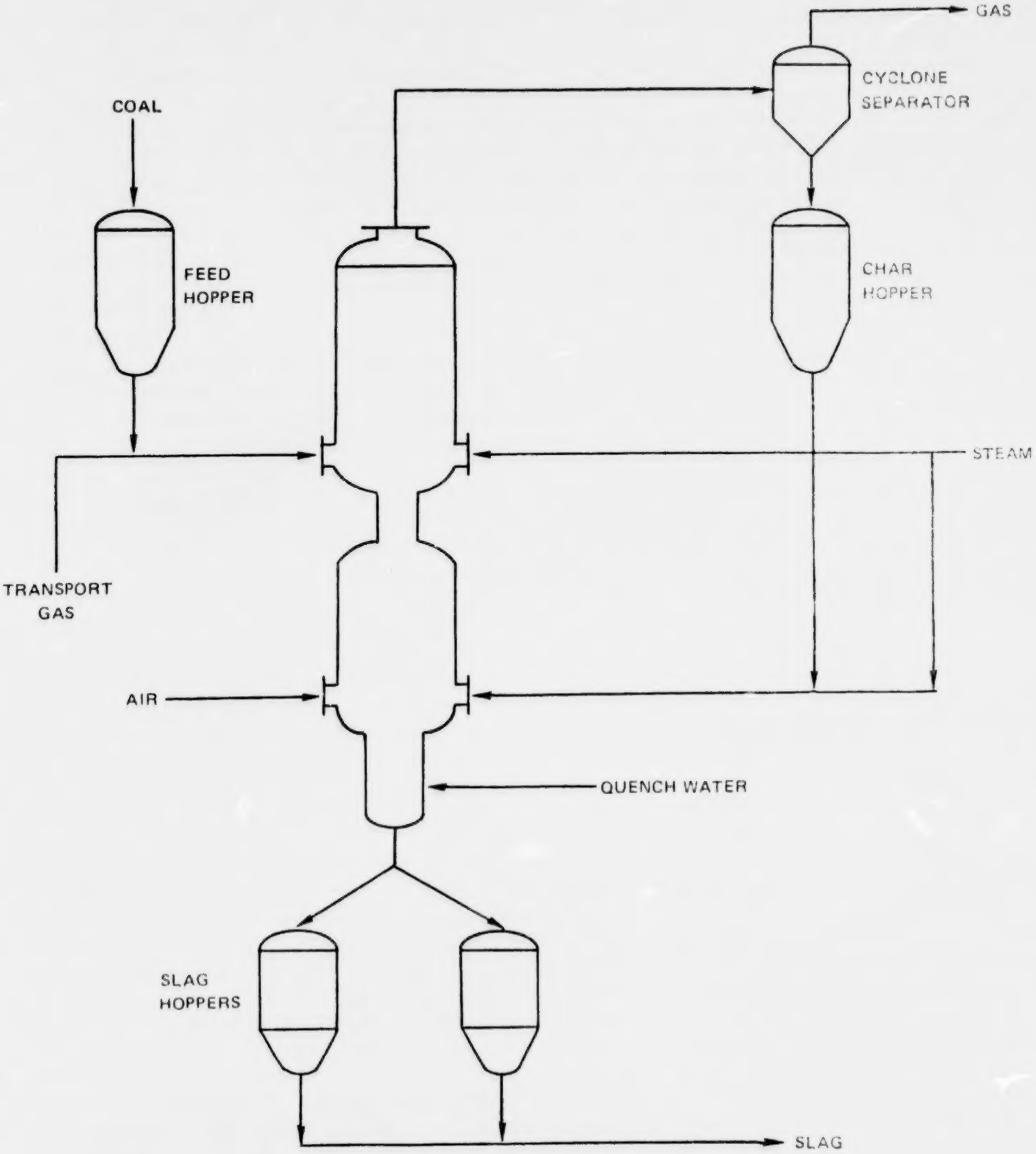
TABLE 1-1 GASIFIER TYPES CONSIDERED

Generic Type	Specific Gasifiers Considered	Status
Fixed Bed	Morgantown Energy Research Center	Ton/hr test unit; other fixed bed designs are commercially available
Entrained Bed	Bituminous Coal Research (BCR) Two Stage	Oxygen-blown unit for syn gas production undergoing tests; air-blown version being designed by Foster Wheeler Energy Corporation
	Koppers Totzek Atmospheric Pressure	Commerical
	Texaco Partial Oxidation	Commercial on oil
Fluid Bed	Institute of Gas Technology U-Gas	Bench scale test
Molten Salt	Pullman Kellogg and Rockwell International	Ton/hr test unit being built by Rockwell

While no practical minimum steam feed rate has been established, on a theoretical basis a value of 0.144 lb steam per lb coal was determined. Previous studies had used values as high as 0.57. This reduction, coupled with one in the coal transport gas flow rate, produced an increase from 78.5 percent to 83 percent in predicted cold gas efficiency (fuel gas chemical heating value/coal chemical heating value). This improved efficiency is accompanied by a reduction in both water vapor and carbon dioxide in the raw fuel gas. By reducing the amount of steam diluent, it is theoretically possible to come close to a 1:1 ratio between oxygen and carbon molecules in the feed, thereby improving the product gas heating value due to the increased production of CO instead of CO_2.

FIG.1–2

BCR ENTRAINED – FLOW GASIFIER


GAS
CYCLONE SEPARATOR
COAL
FEED HOPPER
CHAR HOPPER
STEAM
TRANSPORT GAS
AIR
QUENCH WATER
SLAG HOPPERS
SLAG

Aside from an improved coal to gas conversion efficiency there are several other beneficial effects. Reduced water vapor concentration means less heat is lost at low temperature due to condensation of water in the fuel gas stream. The increased chemical energy can be converted at combined-cycle efficiency whereas sensible energy associated with CO_2 production in the gasifier for the most part can only be converted at steam cycle efficiency. Also, equipment capital savings are realized due to reductions in water purification requirements, condensate processing and fuel gas volume.

The modifications of gasifier operating conditions can also produce changes in the formation of potential pollutants in the gasifier. The sulfur in the feed coal appears mainly as H_2S and COS after gasification. Other sulfur compounds are present but have not been satisfactorily identified. When operating with excess steam, chemical equilibrium calculations for the entrained bed gasifier indicate that approximately 96 percent of the sulfur would appear as H_2S and nearly 4 percent as COS. At the low steam flow rates, the same calculations indicate approximately 6 percent COS. This could affect the size and performance of the sulfur cleanup system as will be described later.

The gasifier operating conditions also could affect the formation of NO_x. The two major sources of NO_x emissions are fuel-bound nitrogen compounds which are converted to NO_x during combustion and thermal NO_x which results from locally high combustion temperatures. Equilibrium calculations for the entrained bed gasifier indicate a 50 percent reduction in fuel-bound nitrogen compounds (taken as ammonia) when the gasifier is operated at the low steam conditions. While essentially all the fuel-bound nitrogen compounds can be removed by water scrubbing as part of the low-temperature gas cleanup system, there has been no method determined for removal at high-temperature.

The previous discussion has focused on a particular type of gasifier, the air-blown entrained bed type. In many respects the other gasifiers studied are similar; however, some significant differences should be noted. Both the fluid-bed (U-Gas) and molten salt types can expect a high degree of catalytic activity as a result of their basic design concept. This should be quite desirable since a major concern in other designs is that COS and NH_3 levels will exceed equilibrium values.

Oxygen, rather than air, has long been used as the oxidant in gasifiers when the desired off gas consists mainly of H_2 and CO with a small amount of CO_2. Usually called 'syn-gas,' this gas with a heating value of approximately 300 Btu/SCF is typically used in the production of H_2. Because of the larger technology base of oxygen-blown gasification and also because of the similarity of its combustion to that of natural gas, there has been considerable interest in the use of this medium-Btu gas in the utility industry as well as in industrial boilers.

In order to ascertain the benefits, if any, that could be obtained from the use of this gas in integrated power plants, a comparison of power plants with oxygen- and air-blown entrained flow gasifiers with both high- and low-temperature cleanup was made. The results indicated that the power plant with the oxygen-blown gasifier had higher heat rates, higher costs and the potential for very high NO_x emissions as compared to its air-blown counterpart.

Fuel Gas Cleanup Systems

The key advantage of the integrated gasified-coal, combined-cycle system is the opportunity to remove sulfur compounds prior to combustion while the gas is at pressure. The process whereby the fuel gas is desulfurized is the focal point of the environmental impact of the system. Both high- and low-temperature cleanup systems have been investigated. Summaries of the various cleanup systems reviewed are given in Table 1-2 (low-temperature) and Table 1-3 (high-temperature). The Selexol* process (Figure 1-3) of Allied Chemical was selected as typical of the low-temperature system while the half-calcined dolomite process (Figure 1-4) of Conoco was selected for the high-temperature system.

Similar to other sulfur removal systems, the Selexol process is sensitive to the type of sulfur compound that must be removed, i.e., whether sulfur in the form of COS must be removed in addition to H_2S. While H_2S is quite soluble compared to other gases in the fuel, COS is only about 30 percent as soluble as H_2S. As a result, if the system must be sized for appreciable COS removal to meet very low levels of sulfur emissions, its cost and utility requirements increase dramatically.

In discussing sulfur emission levels, it is important to note that all sulfur has been assumed to appear as either H_2S or COS in the fuel gas. Thus, while estimates may show emission levels of less than 0.1 $lb/10^6$ Btu, these do not account for other compounds that may not be removed or processed by the particular sulfur removal or recovery system. For example, data for a Synthane gasifier (oxygen-blown) show a total concentration of up to 100 ppm of other sulfur bearing compounds. If none of these were removed, the resultant emission from these compounds alone would be on the order of 0.05 to 0.1 $lb/10^6$ Btu. While data for this type of gasifier are not directly applicable, they do provide an indication of the limitations of this study.

In the design of the Selexol sulfur removal system, operating temperature is rather critical. Gas sweetening plants generally run with subambient absorber temperatures. However, in the interests of simplifying the process for application in the utility industry, a study was made of the effect of Selexol operating temperature on the overall power plant performance and cost. The results of this study indicated a decided reduction in the cost of electricity due to lower solvent flow rate associated with low-temperature operation. It should be noted that this study was based upon a Selexol system designed for virtually complete H_2S removal from the fuel gas (down to

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*A registered trademark of Allied Chemical Corp.

TABLE 1-2

LOW TEMPERATURE CLEANUP PROCESSES

Basis: 8400 tons/day Illinois No. 6 Coal Fed to BCR Gasifier, or 6700 ppm of Influent H_2S

Process	Absorbent	Type of Absorbent	Temp. °F	Pressure	Efficiency of S Removal		Absorbent Characteristics				Form of Sulfur Recovery	Status
					% H_2S Influent	Effluent H_2S ppm	Life	Regeneration	Selectivity toward	Make up rate		
Chemical solvent type												
1. MEA	Monoethanolamine	Aqueous solution	80 to 120	Insensitive to variation in pressure	99	~100		Thermal	Forms non-regen. comp. with COS, CS_2	50 to 100%	As H_2S gas	Commercial
2. DEA	Diethanol amine	Aqueous solution	100 to 130	Insensitive to variation in pressure	99	~100		Thermal	Absorbs CO_2, does not absorb COS, CS_2	< 5%	As H_2S gas	Commercial
3. TEA	Triethanolamine	Aqueous solution	100 to 150	Insensitive to variation in pressure	99	~100		Thermal	H_2S	< 5%	As H_2S gas	Commercial
4. Alkazid	Potassium dimethyl amino acetate	Aqueous solution	70 to 120	Insensitive to variation in pressure 1 - 80 atm	99	~100		With steam	H_2S		As H_2S gas	Commercial
5. Benfield	Activated potassium carbonate solution	Aqueous solution	150 to 250		99	H_2S + COS ~100	Unlimited No degradation	With steam	H_2S is high		As H_2S gas	Commercial
6. Catacarb	Activated potassium carbonate solution	Aqueous solution	150 to 250	Insensitive to variation in pressure generally > 300 psi	99	H_2S + COS ~100		With steam	H_2S - partial also absorbs COS, CS_2	< 5%	As H_2S gas	Commercial
Physical solvent type												
7. Sulfinol	Sulfolane + Dilsopropanoamine	Organic solvent	80 to 120	High pressure preferred	99	H_2S + COS ~100		Low pressure heating or with steam	H_2S, and also absorbs COS, CS_2 and mercaptans		As H_2S gas	Commercial
8. Selexol	Polyethylene glycol ether	Organic solvent	20 to 80		99	H_2S + COS ~100			H_2S also absorbs COS		As H_2S gas	Commercial
9. Rectisol	Methanol	Organic solvent	< 0		99	~100			H_2S			Commercial
Direct conversion												
10. Stretford	Na_2CO_3 + anthraquinone sulfonic acid	Alkaline solution			99.9	~10			H_2S	50 to 100%	Elemental sulfur	Commercial
11. Townsend	Triethylene glycol	Aqueous solution	150 to 250		99.9	~10			H_2S		Elemental sulfur	
Drybed type												
12. Iron sponge	Hydrated Fe_2O_3	Fixed bed	70 to 100		99	H_2S + COS ~100			H_2S and also towards COS, CS_2 and mercaptans		Elemental sulfur	Commercial

TABLE 1-3

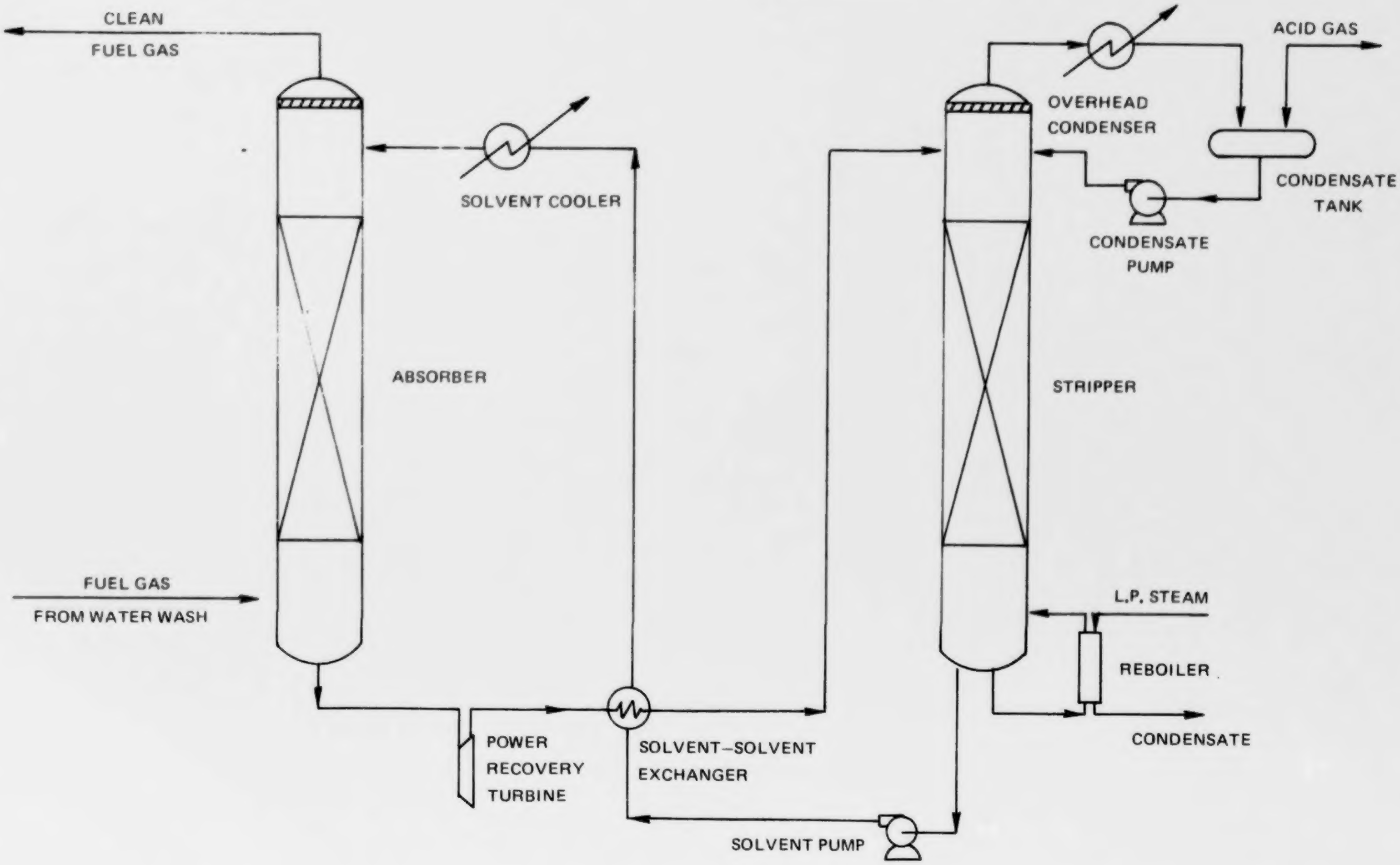
HIGH TEMPERATURE CLEANUP PROCESSES

Basis: 8400 tons/day Illinois No. 6 Coal Fed to BCR Gasifier, or 6700 ppm of Influent H_2S

Process	Absorbent	Type of Bed	Temp. °F	Pressure	Efficiency of S Removal		Absorbent Characteristics				Form of Sulfur Recovery	Energy Required		Status
					%H_2S Influent	Effluent H_2S ppm	Life	Regeneration	Selectivity toward	Make up rate		Elec. kw	Other stu	
1. Bureau of Mines	Sintered pellets of Fe_2O_3 (25%) and fly ash	Fixed bed	1000 to 1500	Insensitive to variation in pressure	~95	~350	>174 cycles Wt loss < 5%	With air	H_2S, COS	<5%	As SO_2 gas			Pilot
2. Babcock and Wilcox	Fe_2O_3	Fixed bed	800 to 1200	Insensitive to variation in pressure	~99	~75					As 12-15% SO_2 gas			Experimental
3. CONOCO	Half calcined dolomite	Fluidized bed	1500 to 1800	~200 psia H_2S removal is high at low pressure	~95	~350		10-13% with steam and CO_2	H_2S, COS	1% of circulation rate	As H_2S gas to Claus process	96.360		Pilot
4. Air products	Calcined dolomite	Fixed bed	1600 to 2000	Insensitive to variation in pressure			minimum 5-6 cycles	80-90% with steam and CO_2	H_2S, COS		As H_2S gas to Claus process			Abandoned
5. Battelle Northwest	Molten carbonates (15% $CaCO_3$)	Solution	1100 to 1700	Atmospheric H_2S removal is high at low pressure, 5-6 psig	~95	~350		With steam and CO_2	H_2S, COS, fly ash		As H_2S gas to Claus process			Pilot
6. IGT - Meissner	Molten metal (proprietary)	Splashing contact	900		~98	~150		Electrolytic	H_2S, COS			9830		Conceptual

FIG.1–3

SELEXOL LOW – TEMPERATURE DESULFURIZATION


CLEAN
FUEL GAS
SOLVENT COOLER
ABSORBER
FUEL GAS
FROM WATER WASH
POWER
RECOVERY
TURBINE
SOLVENT–SOLVENT
EXCHANGER
SOLVENT PUMP
OVERHEAD
CONDENSER
ACID GAS
CONDENSATE
TANK
CONDENSATE
PUMP
STRIPPER
L.P. STEAM
REBOILER
CONDENSATE

77–08–299–2

CONOCO HIGH – TEMPERATURE DESULFURIZATION

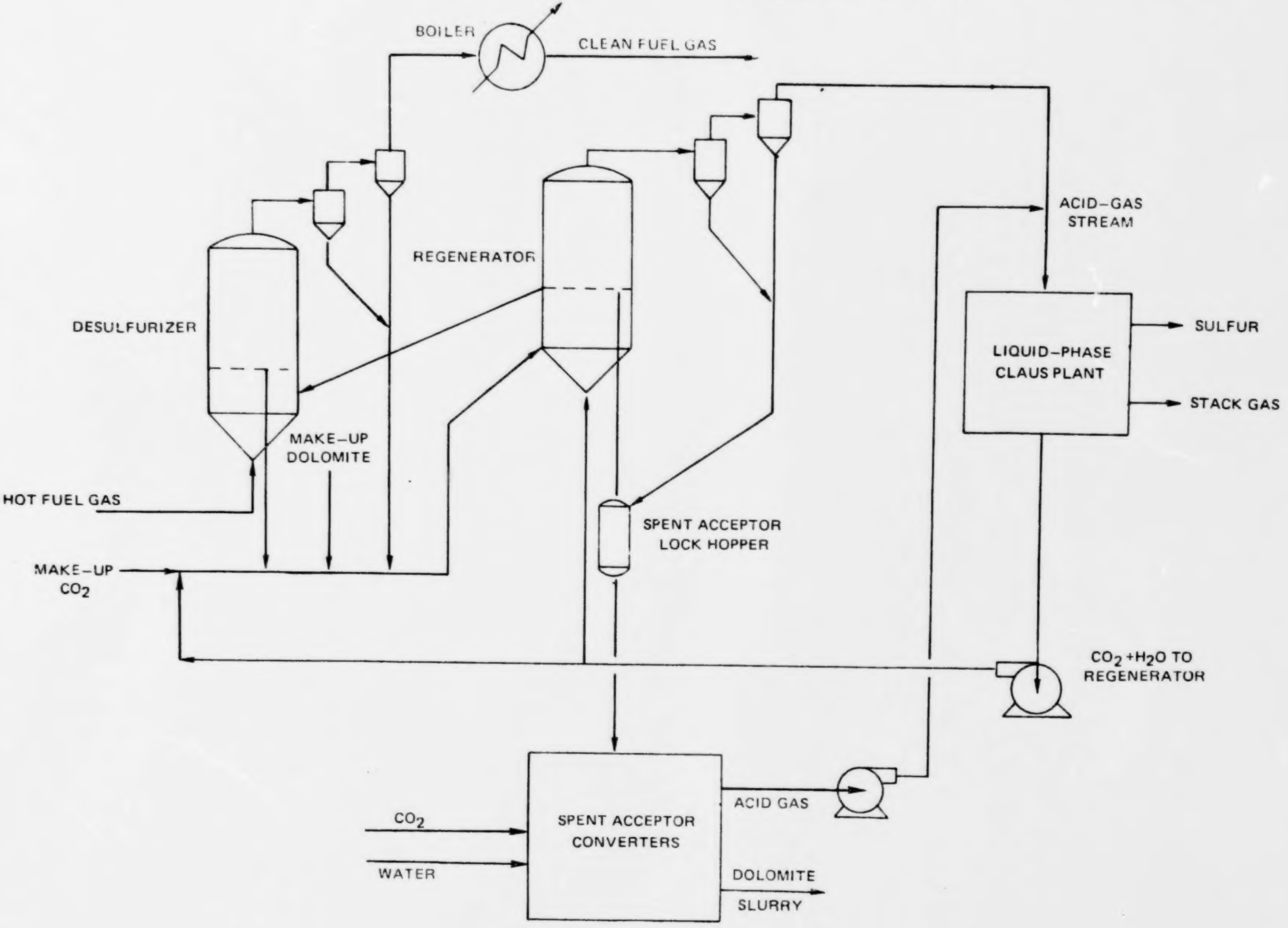

BOILER
CLEAN FUEL GAS
REGENERATOR
DESULFURIZER
ACID–GAS STREAM
LIQUID–PHASE CLAUS PLANT
SULFUR
STACK GAS
MAKE–UP DOLOMITE
HOT FUEL GAS
MAKE–UP CO_2
SPENT ACCEPTOR LOCK HOPPER
CO_2 + H_2O TO REGENERATOR
CO_2
WATER
SPENT ACCEPTOR CONVERTERS
ACID GAS
DOLOMITE SLURRY

FIG.1–4

approximately 35 ppm) and that total power plant sulfur emissions were on the order of 0.2 to 0.4 lb/10^6 Btu. If less ambitious sulfur removal goals were considered (Reference 1-7), it appears that there would be a smaller difference between ambient and refrigerated operation.

Performance was estimated for power plants using each of the gasifier types followed by the Conoco half-calcined dolomite high-temperature desulfurization system. Reactions within the desulfurizer are reported to be close to equilibrium and are limited by the amount of water vapor and carbon dioxide in the fuel gas. Low concentrations of these molecules in the raw fuel gas improve desulfurization. The latest data for the air-blown entrained bed gasifier with minimum steam feed were especially attractive for use in a combined-cycle application with the high-temperature desulfurizer. To maintain gasifier temperature within design limits under minimum steam feed, oxidant feed must also be low; hence, the product gas has low concentrations of both water vapor and carbon dioxide. Desulfurizer performance under these conditions resulted in an estimated emission after combustion of 0.09 lb $SO_2/10^6$ Btu. Similarly, use of the gasifier equilibrium model to simulate the fluid bed/high-temperature cleanup system at reduced steam feed rates indicated emissions of approximately 0.15 lb $SO_2/10^6$ Btu.

Removal of particulates and nitrogen compounds also must be considered when evaluating fuel cleanup processes. For example, the level of particulates needed to avoid excessive high-temperature turbine erosion is lower than the environmental regulations. Thus, the water wash associated with low-temperature systems which clean the fuel gas stream to gas turbine tolerance will meet the EPA particulate standard. No such system is available for high-temperature processes.

Fuel cleanup system characteristics also affect the quantity of nitrogen compounds in the fuel which could form NO_x during combustion. For systems using low-temperature sulfur removal, a water scrub is incorporated which effectively removes both particulates and ammonia, the major nitrogen compound after gasification. Removing ammonia minimizes NO_x formed during combustion in conventional burners but thermally generated NO_x is still a problem. However, United Technologies sponsored tests carried out in conjunction with the Texaco Development Company at their Montebello California Research Laboratory have led to the identification of a potentially attractive burner concept. The test results (References 1-8 and 1-9) indicate that a premix burner should meet both current and proposed NO_x standards. Thorough premixing of fuel and air permits control of combustion temperature and associated NO production. To avoid possible preignition, fuel temperature to the gas turbine has been limited to 1000 F with both high- and low-temperature cleanup systems. While this in effect penalizes the high-temperature cleanup system by not allowing it to utilize the fuel gas sensible heat over 1000 F in the combined cycle, this penalty is necessary if excessive NO_x emissions are to be avoided.

Power System

Throughout the course of the ongoing studies, the power system has been updated to reflect advances in gas turbine technology. The latest update includes results from the recently completed DOE-sponsored High-Temperature Turbine Technology Program (HTTTP) Phase I. The advanced turbine defined in this program has a dual-spool gas generator and a free turbine driving the generator. The overall pressure ratio is 18:1 at a turbine inlet temperature of 2600 F. Compressor discharge air is precooled to 400 F prior to use in cooling the rotating blades while water cooling is used in the static sections. Advanced ceramic coatings were projected for a number of hot section static parts. The characteristics of the power system are given in Table 1-4.

The present study has adapted the flow path of the HTTTP engine but has assumed ceramic vanes, which require no cooling, rather than the water cooled concept. The resulting performance is essentially the same.

A gas turbine pressure ratio of 18:1 results in an exhaust temperature that is sufficiently high to provide both superheat and reheating for a 2400 psi, 950 F reheat steam cycle. The use of such a high-performance steam system would not be practical in a conventional distillate-fueled combined cycle. This is one of the advantages of the fully integrated system. Little, if any, evaporation is done in the heat recovery boiler in the gas turbine exhaust stream. High-pressure saturated steam is raised in the gasifier and cleanup process and can be extracted from the steam cycle for process use after having produced some useful output. The net effect is a range of overall coal to power conversion efficiencies from 42 percent to 46 percent. These high efficiencies are important not only because they minimize the fuel dependent cost of power generation but also because increased plant efficiency implies increased electrical output for a given size coal processing system. Thus, the capital cost of fuel processing equipment per unit output is reduced. This, in turn, affects the degree to which the fuel can be cleaned within existing economic constraints.

OVERALL POWER PLANT

The performance, cost, and environmental intrusion of the power plant cannot be defined until the various components have been integrated. The integrated plant is very complex involving the exchange of air and steam between the fuel processing and power systems. A simplified schematic of a typical integrated power plant with low-temperature desulfurization is shown in Figure 1-5.

Performance

A summary of the performance, costs and major air emissions of eight integrated power plants is given in Table 1-5. Integrated power plants using fixed-bed gasification and the first-generation of advanced gas turbines

TABLE 1-4

POWER SYSTEM CHARACTERISTICS(1)

Gas Turbine	
Inlet Temperature, F	2600
Pressure Ratio	18
Inlet Flow Rate, lb/sec	983
Unit Power, MW	181
Steam Turbine	
Throttle/Reheat Temperature, F	950/950
Throttle/Reheat Pressure, Psig	2400/585
Throttle/Reheat Flow Rate, lb/sec	764/764
Condenser Pressure, in Hg	4
Unit Power, MW	453
Overall Power Plant	
Gross Output, MW	1178
Net Output, MW	1098
Heat Rate,(2) Btu/kWh	7775

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(1) For integrated plants with air-blown entrained bed gasifiers and low-temperature fuel gas cleanup. Other types of gasifiers and cleanup systems will affect power system characteristics.

(2) Coal pile to busbar.

FIG.1–5

BCR/SELEXOL SYSTEM

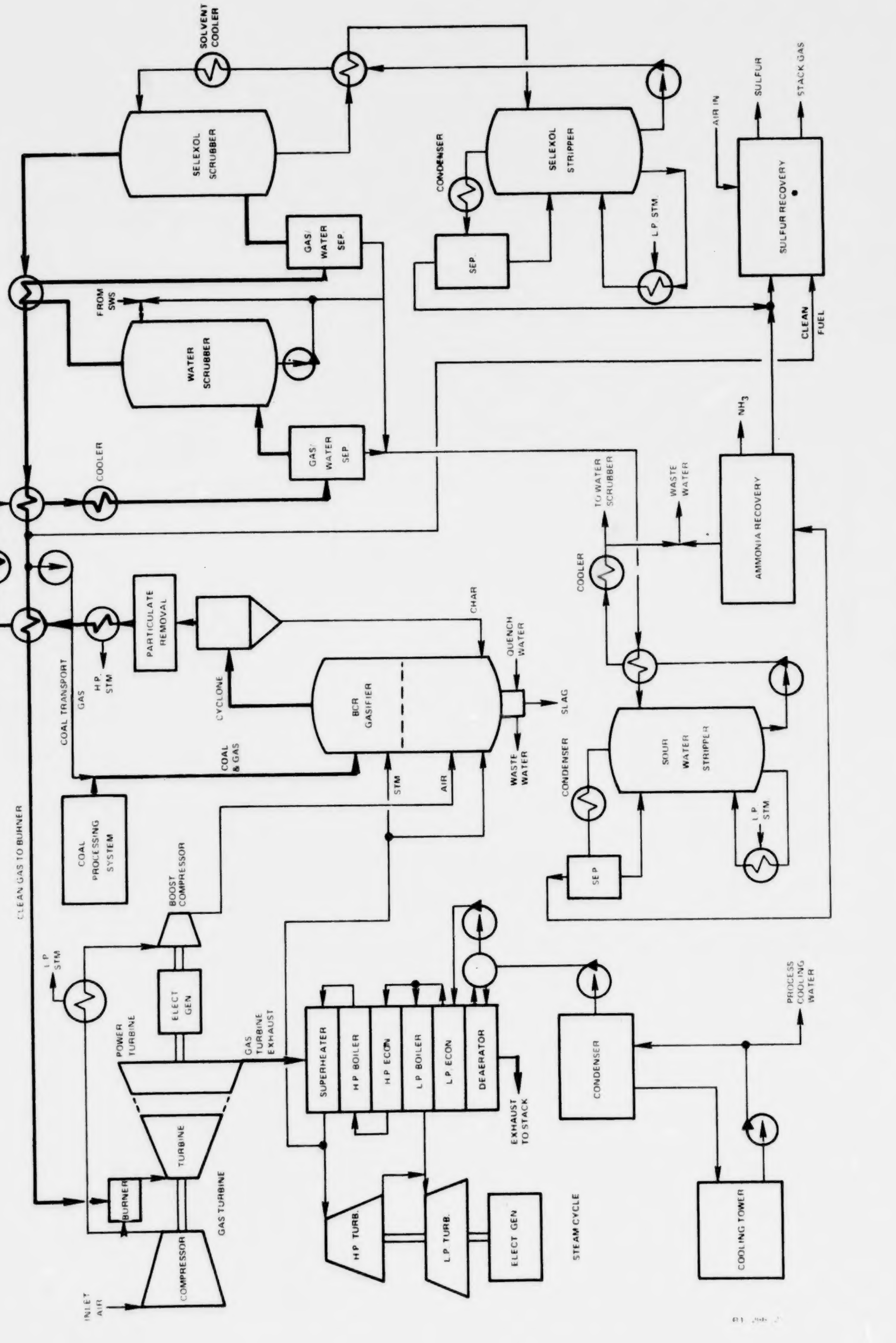

SOLVENT COOLER
SELEXOL SCRUBBER
GAS/ WATER SEP.
FROM SWS
WATER SCRUBBER
GAS/ WATER SEP.
COOLER
L P STM
CONDENSER
SELEXOL STRIPPER
SEP
L P STM
AIR IN
SULFUR
STACK GAS
SULFUR RECOVERY
CLEAN FUEL
NH$_3$
TO WATER SCRUBBER
WASTE WATER
AMMONIA RECOVERY
COOLER
PARTICULATE REMOVAL
H.P. STM
CHAR
QUENCH WATER
CYCLONE
BCR GASIFIER
SLAG
COAL TRANSPORT GAS
COAL & GAS
STM
AIR
WASTE WATER
CONDENSER
SOUR WATER STRIPPER
L P STM
SEP
CLEAN GAS TO BURNER
COAL PROCESSING SYSTEM
BOOST COMPRESSOR
L P STM
ELECT GEN
POWER TURBINE
GAS TURBINE EXHAUST
SUPERHEATER
H P BOILER
H P ECON
L P BOILER
L P ECON
DEAERATOR
EXHAUST TO STACK
CONDENSER
PROCESS COOLING WATER
INLET AIR
BURNER
TURBINE
COMPRESSOR
GAS TURBINE
H P TURB.
L P TURB.
ELECT GEN
STEAM CYCLE
COOLING TOWER

TABLE 1-5

SUMMARY OF POWER PLANT CHARACTERISTICS

Powerplant Type	Air-Blown BCR/ Selexol	Air-Blown U-Gas/ Selexol(1)	Air-Blown U-Gas/ Selexol(2)	Air-Blown BCR/ Conoco	Air-Blown U-Gas/ Conoco(2)	Air-Blown Molten Salt	O_2-Blown BCR/ Selexol	O_2-Blown BCR/ Conoco
Net Power(3) - MW	1098	1074	1109	1136	1145	955	1045	1098
Net Efficiency - %	43.9	42.9	44.3	45.4	45.7	43.3	41.8	43.9
Capital Cost - $/kW	657	663	641	597	543	683	725	709
Power Cost(4) - Mills/kWh	31.76	32.24	31.08	29.25	27.08	32.99	34.87	33.88
lb $SO_2/10^6$ Btu(5)	0.40	0.19	0.23	0.13	0.19	0.21	0.39	0.73
lb $NO_x/10^6$ Btu(6)	--(7)	--(7)	--(7)	2.52(8)	0.165(8)	--(7)	--(7)	2.60(8)
lb particulate/10^6 Btu(6)	<0.01	<0.03	<0.03	<0.10	<0.10	<0.03	<0.03	<0.10

(1) Based on EPRI-furnished data
(2) Based on UTRC/IGT-developed low steam process
(3) Based on 700,000 lb/hr coal flow
(4) Based on $1.00/$10^6$ Btu coal
(5) Emissions from power plant
(6) Emissions from power system
(7) Function of burner design <0.35 lb/10^6 Btu
(8) Fuel bound nitrogen only - burner design will add thermal NO_x

(turbine inlet temperature of 2200 F) have not been included because their performance and costs are significantly less attractive than the more advanced second-generation gasifiers and gas turbines. A detailed discussion of the less-advanced power plants is given in Reference 1-5.

Performance of the various advanced air-blown systems with low-temperature cleanup showed little difference between gasifier types. Generating efficiency from coal pile to busbar varied between 43 and 45 percent. The same is true of the air-blown fluid bed and entrained flow-type gasifiers when coupled with the Conoco high-temperature cleanup system. For these, the performance differences were so small that both can be considered to have an efficiency of 46 percent. Thus, on the basis of performance, there is little difference between integrated plants with either high- or low-temperature cleanup especially when the low-temperature cleanup components are carefully tailored to the power generating system.

It is of interest to note the comparison of integrated power plants with air- vs. oxygen-blown operation of the entrained bed-type gasifier. With low-temperature sulfur removal, a performance difference of approximately two points was estimated (air-blown, $\eta = 0.439$ and oxygen-blown, $\eta = 0.418$). The factors that cause this difference are associated with the loss of mass in the fuel gas due to the oxygen separation process and the relatively high steam feed needed in the oxygen-blown gasifier. In the case of the power plant with the low-temperature cleanup system, the heat of vaporization of the steam is completely lost when the water vapor in the fuel gas is condensed before entering the cleanup system. In the case of high-temperature cleanup, the water vapor remains in the fuel gas and acts in a manner similar to nitrogen as a diluent. While not enhancing performance, the fact that the water vapor is available to do work in the gas turbine, and to provide sensible heat in the waste heat boiler, results in a lesser difference between the oxygen- and air-blown power plants when used with a high-temperature cleanup system. Here the performance is separated by just over one point (air-blown, $\eta = 0.458$ and oxygen blown, $\eta = 0.444$).

Costs

For the air-blown power plants with the Selexol cleanup process, capital costs were quite close for the entrained bed, fluid bed and low-steam version of the fluid bed gasifier, varying from \$630/kW to \$660/kW. The estimated cost of electricity was also virtually the same for each, varying between 31 and 32 mills/kWh. Capital cost of the power plant with the oxygen-blown entrained bed-type system with Selexol cleanup was approximately \$720/kW. While performance of this power plant was estimated to be some two points poorer than that of the air-blown system, the cost of electricity (based on coal at $\$1/10^6$ Btu) was approximately 35 mills/kWh. This small difference of approximately 10 percent, while likely to be real, is well within the estimated accuracy limits of the costs available for use in the study and probably would not represent a real deterrent to the use of an oxygen-blown gasifier should the technology be preferred to that for an air-blown gasifier.

However, considering the use of fuel gas for power generation only, there appears to be no reason to opt for the oxygen-blown gasifier as it shows no advantage in terms of emissions, performance or cost and would clearly be somewhat more complex than the air-blown device.

The molten salt system can be considered to be a low-temperature system since it incorporates a water wash (albeit at 175 F). It was estimated to be comparable in cost to the others although the available cost estimates are less definitive. Capital cost is estimated to be $720/kW while electricity is estimated to cost 33 mills/kWh.

The integrated power plants using high-temperature desulfurization processes show some advantage in terms of reduced cost of electricity. However, they do not include provision for nitrogen removal from the fuel gas and, thus, would not be environmentally acceptable. A reduction of 2 to 3 mills/kW over air-blown Selexol-based systems is indicated, but this advantage is based on costs that include a larger number of unknown factors than do those for the Selexol-based systems. Both cleanup and particulate removal costs are based on extrapolation of small-scale lab test data as opposed to low-temperature equipment which, while not having handled gasifier effluents, has been built and run on a commercial scale. For the power plants using air-blown gasifiers with Conoco cleanup, estimated power costs range from 27 to 29 mills/kWh. The high-temperature, entrained bed, oxygen-blown system has only a slightly lower cost of electricity than that for its low-temperature counterpart (34 vs 35 mills/kWh). Thus, for a power plant with an oxygen-blown gasifier, there appears to be little incentive for the use of a high temperature sulfur removal device. Those plants using air-blown gasifiers show a 2 to 3 mill/kWh improvement and also have the potential for a significant reduction in sulfur emissions. It must be reiterated, however, that the high-temperature systems do not have identifiable methods of removing fuel-bound nitrogen nor have methods of particulate removal been satisfactorily defined.

During the course of this study improvements in integration as well as improvements in both gasifier and power system performance have resulted in an increase in the absolute value of overall power generating efficiency of from 3 to 6 percentage points. This not only means a lower fuel cost, but also permits the capital cost of gasification and cleanup to be spread over a larger power output thus reducing the specific cost of these parts of the system.

Environmental Intrusion

All of the power plants considered easily were able to meet the current EPA standards for stationary power plant sulfur emissions of 1.2 lb $SO_2/10^6$ Btu. The power plant with the oxygen-blown gasifier shows an advantage in terms of sulfur removal system size. Because there is no nitrogen diluent, the quantity of gas to be processed is reduced by more than one-third. With low-temperature cleanup, this results in an overall reduction in system cost and utilities without a large effect on sulfur emissions. For the high-temperature case, equipment size, other than that of the absorber, is more dependent on sulfur loading than gas flow rate and there is little difference in size and cost. However, as noted earlier, due to the much lower water vapor and CO_2 concentrations, the sulfur emissions estimated for the air-blown system are significantly less than for the oxygen-blown system (0.1 vs. 0.7 $lb/10^6$ Btu.

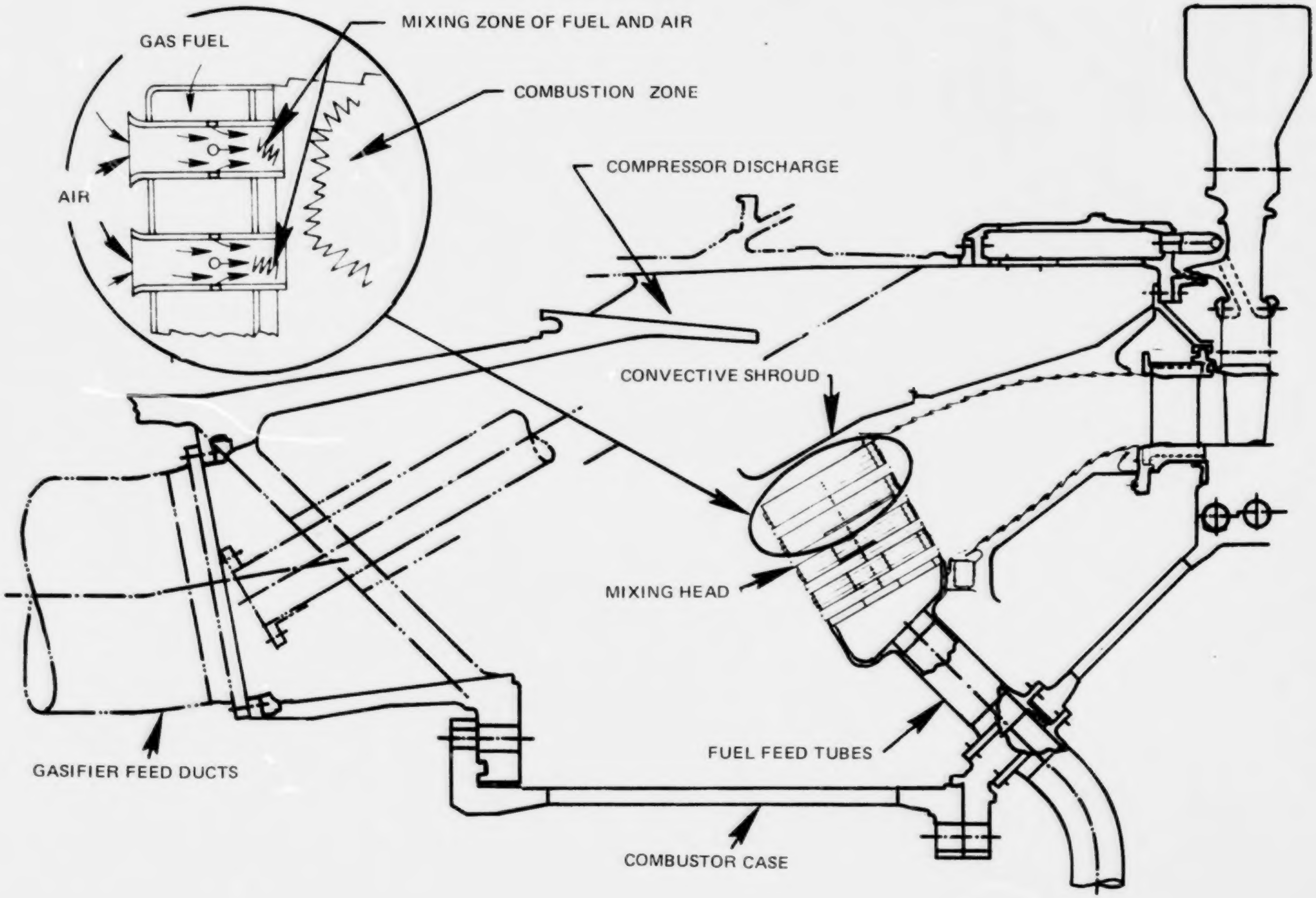
Only the low-temperature systems such as Selexol incorporate a water scrubber for ammonia removal. Thus, in the high-temperature systems, production of NO_x in the gas turbine could be quite high, as much as 2.5 lb $NO_x/10^6$ Btu if all the nitrogen went to the oxide form. (Currently proposed NO_x standards for gas turbines are between 0.35 and 0.9 $lb/10^6$ Btu depending on the level of nitrogen in the fuel. The power plant standard is 0.7 lb $NO_x/10^6$ Btu). Clearly, if the high-temperature cleanup system is to be viable, the combustor design must be such as to convert ammonia and other fuel bound nitrogen compounds to N_2 rather than NO_x. This must be done while limiting the formation of thermal NO_x. Unfortunately, from the viewpoint of thermal NO_x formation, the fuel gas from the oxygen-blown systems has a much higher flame temperature than even methane. Design of a NO_x-free combustor for that fuel could be a problem and certainly will be more difficult than for the low-Btu gas.

For those power plants with the low-temperature cleanup, ammonia has been removed and only thermally generated NO_x is a problem. For thermal NO_x control, several promising combustion techniques have been identified. After regeneration, the hot fuel gas and compressor discharge air combine to produce flame temperatures that would result in excessive NO_x emissions in a conventionally designed combustor. Premixed, fuel-lean mixtures effectively reduce flame temperature and consequently limit thermal production of nitrogen oxides. In the premix concept, shown in Figure 1-6, air and fuel are well mixed prior to combustion thereby significantly reducing the combustion time and temperature and, thus, the production of NO_x. Fuel-lean conditions, however, have been shown to result in the conversion of virtually all ammonia to NO_x, which is unacceptable from an emission standpoint. Research has shown that fuel-rich combustion can convert virtually all of the fuel-bound nitrogen to N_2 while minimizing the production of thermal NO_x. Work currently underway for the EPA at the Government Products Division, Pratt & Whitney Aircraft has demonstrated NO_x emissions as low as 50 ppm (~0.23 $lb/10^6$ Btu) while burning a fuel doped with 0.5 percent nitrogen (Reference 1-10).

In addition to sulfur and nitrogen oxides, sources of emissions exist throughout the coal and fuel gas processing train. Many of these are similar to those in coal-fired steam plants, particularly in regard to coal preparation and heat rejection. Residuals from the cleanup process will include waste water which must be treated prior to disposal. In general, treatment will be highly dependent on individual plant design and will be affected by plant location.

FIG.1–6

POTENTIAL PRE–MIX COMBUSTOR LAYOUT


MIXING ZONE OF FUEL AND AIR
GAS FUEL
COMBUSTION ZONE
AIR
COMPRESSOR DISCHARGE
CONVECTIVE SHROUD
MIXING HEAD
GASIFIER FEED DUCTS
FUEL FEED TUBES
COMBUSTOR CASE

77–09–15–2

The overall environmental intrusion must also consider the many trace elements found in coal. Consideration of trace elements became a significant part of the program effort. Because of the many trace elements present in coal and the difficulty of discussing all of them, a ranking method was applied to limit the number considered. The ranking procedure was based on the ratios of concentration in typical Illinois No. 6 coal to allowable concentration in air, water or solid waste as given by Minimum Acute Toxicity Effluent (MATE) values and Estimated Permissible Concentrations (EPC) for each of the elements. The elements having the highest values of these ratios could have the highest potential for producing harmful effects. This selection process was tempered by consideration of the potential form that the element or its compounds might take, as well as experience as to which have proven to be troublesome in other applications. The elements As, B, Be, Cd, Cr, Hg, Ni, Pb and V were selected as potentially troublesome.

An attempt to discuss the fate of trace elements during the gasification of typical Illinois No. 6 coal and subsequent use of the gas is difficult. From the time that coal enters the plant boundary, there is the potential for it to cause some type of environmental intrusion. Except for some chemicals introduced for water treatment, the coal feed is the source of all trace element emissions. The gasifier itself, a closed vessel, discharges no pollutants directly to the environment. However, coal preparation and feed equipment as well as gasifier exit streams all must be considered.

Those dust and particulate emissions that occur prior to gasification will likely contain trace elements in a distribution of concentration similar to that of the parent coal. The fate of trace elements in the gasifier, whether they leave in the ash/slag or off-gas, is a function of factors such as gasifier configuration, bed type and operating conditions. Trace elements by themselves are not very volatile. Yet slag or ash particles from gasifiers seem to be significantly depleted in most of the trace elements in coal. In the high-temperature reducing atmospheres typical of gasifiers, trace elements thus, would seem to be volatilized in the form of compounds such as carbonyls, sulfides and hydrides. Those compounds condense in quench water.

While there are few data available on the fate of trace elements, a number of estimates have been based on data from Reference 1-11. Table 1-6 presents the results of a calculation based on such an estimate to determine the distribution of the elements between the slag and sour water. It is based on the assumption that the amount of each element not appearing in the ash will be found in the sour water. It shows that mercury, arsenic and cadmium go predominantly to the sour water and may require special care. It may be possible to utilize the sour water in the ash quenching process. If all the elements were to appear in the slag or ash, disposal might be simplified if guards against leaching are taken. As an example of the potential magnitude of the problem, should all of the nickel estimated to appear in the sour water be allowed to escape, it would require 3.6 million gallons per minute of water to dilute that element to the level of the EPC.

TABLE 1-6 POTENTIAL FATE OF TRACE ELEMENTS*

Element	Feed	Flowrate (lb/hr) Final Residue	Sour Water	% in Residue/ % in Sour Water
Hg	0.084	0.008	0.076	10/90
As	16.8	4.4	12.4	26/74
Pb	7.7	4.9	2.8	64/36
Cd	0.62	0.23	0.39	27/63
V	11.9	8.3	3.6	70/30
Ni	10.5	8.0	2.5	76/24
Be	0.70	0.57	0.13	81/19
B	140	126	14	90/10
Cr	10.5	10.5	0	100/0

*Based on a coal feed rate of 700,000 lb/hr

CONCLUDING REMARKS

The results of this study have updated prior work and have further strengthened the conclusion that the integrated gasification/sulfur cleanup/ combined-cycle power system will be one of the more environmentally and economically attractive medthods of generating electric power in the future.

Further work in the areas of the costs and performance penalities associated with deep sulfur removal (removal of significant amounts of COS) appears necessary. Also, it is important to recognize that an evaluation must be made of methods to remove fuel-bound nitrogen prior to combustion and of possible combustor modifications which might convert these compounds to N_2.

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SECTION 2

INTRODUCTION

Previous EPA-sponsored studies (References 2-1 and 2-2) at UTRC demonstrated the potential environmental and economic benefits of integrated coal gasification/gas cleanup/combined-cycle power plants. However, there remained several technical problem areas requiring investigation or further definition. One such area involved an expanded study of plant effluents including an examination of the possible fate of potentially hazardous trace elements. Another problem area was the comparison of integrated power plants using air-blown versus oxygen-blown gasifiers supplying fuel to the advanced power systems. To date, neither have been compared on a consistent basis using the latest data on gasifier performance and effluent discharges. A third area involved the use of advanced technology gasifiers such as the fluid bed and molten salt types which have the potential for higher integrated power plant performance, but which have not been well defined in the integrated power system mode from the technical, economic and environmental viewpoint. Also, to provide a better understanding of the basic capabilities and limitations of these advanced gasifiers, the modeling work done in the previous program phases needed to be extended to consider them. The current program addresses these areas in a manner consistent with the earlier work to provide a meaningful comparison of these gasifiers on both an environmental and economic basis when operating in an integrated power generating system using coal gasification and advanced combined gas and steam turbine cycles with fuel gas cleanup before combustion.

Prior investigations carried out under EPA Contract No. 68-02-1099 resulted in the major conclusion that integrated power plants using low-Btu coal gasification, fuel gas sulfur removal, and combined gas and steam cycle power generation offer the potential for reduced environmental intrusion while generating electric power at costs competitive with or less than conventional coal-fired stations with flue gas desulfurization. Coal processing systems studied in UTRC's prior work for the EPA included entrained-flow (BCR-type) and fixed bed (Morgantown Energy Reseach Center design, referred to herein as the BuMines gasifier) medium pressure (25-atm) gasifiers with both high- and low-temperature cleanup systems.

Each of the coal conversion processes was mated with a combined-cycle generating system for producing electric power. Additionally, systems utilizing existing technology such as the Koppers-Totzek atmospheric-pressure, oxygen blown coal gasifier, and the Shell or Texaco-type partial oxidation oil gasifier were studied. These studies are reported in References 2-1 and 2-2. For continuity, the results are summarized in the early sections of this report.

A further study consideration was the utilization of the latest data developed under the sponsorship of other agencies or by other contractors that could affect the conclusions of these previous studies. For example, gasifier modeling performed during the earlier EPA-sponsored studies had shown the advantages of low steam feed rates. Heat and mass balances available at that time were not optimized for integration with combined-cycle power generating systems. As yet unpublished data, made available by the Electric Power Research Institute (EPRI), substantiated that a marked performance improvement can be achieved by that relatively simple change in approach. Other programs, such as the High Temperature Turbine Technology Program sponsored by DOE have provided technological data regarding the potential for improved combined-cycle performance. Results from that and other ongoing programs have also served as a basis for the selection of the most economically attractive combined-cycle configuration consistent with current and predicted future coal prices.

Because of the importance of these issues to the electric power generating industry, EPRI has participated in this program by allowing the use of information generated under contracts sponsored directly by them. In substance, EPRI has provided process data on air- and oxygen-blown two-stage gasifiers and on advanced fluid-bed gasifiers. Many of these data were developed by Fluor Engineers and Constructors as part of an EPRI-sponsored study. That study, subsequently reported on in Reference 2-3, sought to identify whether significant economic and/or environmental incentives exist for using various current and advanced gasification processes coupled with combined-cycle power plants to produce electricity. The study did not explore in depth the environmental intrusion of these integrated power plants. However, it provided the latest available revisions of the respective gasifier operating conditions.

Fluor has also acted as a direct subcontractor to UTRC to provide process engineering in the areas of fuel gas cleanup, system integration, and effluent definition and control. Hittman Associates, who participated in the earlier study, has continued as a subcontractor to provide overall evaluation of the various discharges of air and water pollutants and solid wastes from the integrated plant. Allied Chemical Co. has provided performance and cost estimates for the Selexol desulfurization system and the Institute of Gas Technology has cooperated in studies aimed at reduced steam feed to the U-Gas gasifier. UTRC has provided power system characteristics and analysis, overall systems integration, and environmental intrusions.

In order to achieve the goal of comparing the environmental, performance and cost characteristics of integrated power plants having both low- and high-temperature sulfur cleanup processes, five program objectives were defined:

1. To provide a more detailed evaluation than was done in previous studies of the effluents from the integrated power plant and the control methods that could be applied.
2. To compare the operational, economic, and environmental characteristics of integrated power plants using oxygen vs. air-blown gasification in an entrained flow (BCR-type) gasifier with both high- and low-temperature cleanup.
3. To consider the merits of advanced gasifiers such as the fluid bed and the molten salt gasifiers when integrated with a combined-cycle power system.
4. To update previous work so that a meaningful comparison of performance, emissions, and cost can be made between the present and previous systems.
5. To utilize modeling techniques developed under previous contracts to identify the effect of gasifier operating parameters on overall system emissions and performance characteristics.

In carrying out these objectives, several other areas were uncovered that warranted investigation. Among these were: 1) the effect of operating temperature on the Selexol system performance and cost; 2) the performance of the fluid-bed gasifier at reduced steam feed rates; and 3) the use of nonselective, low-temperature desulfurization systems.

The conclusions from the study and recommendations for further work follow this introduction. Sections 5, 6 and 7 deal with the individual processes which make up the integrated gasification/sulfur cleanup/combined-cycle power plant. Subsequent sections describe the overall power plant integration, power plant air emissions and other environmental intrusion. A final section presents the performance and resultant cost of electricity for the integrated power plants.

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SECTION 3

CONCLUSIONS

1. Revisions in gasifier and power system operational characteristics have resulted in lower integrated power plant heat rate and reduced emissions compared to the estimates made for similar integrated power plants in the prior UTRC studies carried out for the EPA.

2. The revised integrated power plants have become an even more attractive method for providing electricity at a lower cost with less environmental intrusion than a conventional steam plant with flue gas desulfurization.

3. Integrated power plants having oxygen-blown gasifiers with either low- or high-temperature sulfur cleanup appear to be less attractive than their air-blown counterparts. They exhibit higher heat rates, higher costs and the potential for higher emissions of NO_x.

4. For the integrated power plants studied having various advanced, air-blown gasification systems, little difference between power plants was noted in emissions, performance or generating cost when operating with low-temperature desulfurization and at low gasifier steam feed rates. Thus, gasifier selection may be based on other considerations.

5. Power plants with a high-temperature cleanup process, for which bench scale data only are available, have estimated performance and cost advantages over plants with low-temperature sulfur cleanup, for which process data are available. Careful design of plants with low-temperature cleanup can reduce these estimated advantages to marginal levels.

6. In high-temperature cleanup processes, the problems associated with efficient removal of particulates less than 10 μ and the removal or harmless conversion of nitrogen compounds in the fuel gas remain to be resolved. Until such systems can be better identified, associated cost and performance penalties cannot be accurately defined.

7. At the elevated fuel gas temperatures needed to attain highly efficient, integrated power plant operation, conventional gas turbine combustor designs would result in excessive NO_x formation even with low-Btu gas. A premix burner offers the potential for off-stoichiometric combustion and greatly reduced NO_x.

8. For the selected low-temperature desulfurization system (Selexol), removal of sulfur to very low levels can double equipment cost if it is necessary to design for COS removal.

9. Any advantages in power plant performance or cost resulting from the use of an ambient temperature Selexol process rather than a refrigerated Selexol process are lost when sulfur removal below the level of 1.2 lb $SO_2/10^6$ Btu is desired.

10. Data needed to predict the fate of the numerous trace elements in the coal feed are scarce. While not definitive, it appears that water scrubbing of the fuel gas would be desirable in controlling the discharge of the compounds of these elements.

SECTION 4

RECOMMENDATIONS FOR FURTHER WORK

1. Further investigations are required to identify the cost and effect on overall power plant performance of sulfur removal to levels of 95 percent and beyond.

2. An investigation should be conducted to compare the effects of removing fuel-bound nitrogen prior to combustion with the effects of combustion modifications. Of interest are the amounts of nitrogen removed, the cost, effect on overall system performance and the reduction in nitrogen oxide emissions.

SECTION 5

OVERVIEW OF GASIFICATION AND CLEANUP PROCESSES FOR USE WITH COMBINED-CYCLE POWER PLANTS

INTRODUCTION

The basic technology for production of low-Btu gas is available. Gasification has been in common use since the 1800's. A number of desulfurization processes have been developed for gas cleanup. These are designed for use in sweetening natural gas as well as in the cleanup of the various types of low-Btu or synthesis gas.

In addition to these existing processes, a number of advanced types are in various stages of development. In general, they are intended to provide for more efficient gasification of widely varying coal types while having increased throughput and consequently lower cost. This section presents a brief description of the basic types of gasifiers and cleanup processes that are potentially available and evaluates their applicability for use in combined cycle power generation.

System Arrangement

An integrated system is shown schematically in Figure 5-1. It shows the primary interconnections between the gas turbine, steam bottoming cycle and coal processing or gasification and cleanup system. The gasification process requires air which can be supplied from the compressor discharge of the gas turbine. Using this high pressure air permits pressurized operation of the entire fuel processing system. This is generally beneficial since it reduces component size and improves the performance of many desulfurization systems. Also, many gasification processes need steam which can be supplied by extraction from the steam cycle after having done some useful work. Low pressure steam can also be extracted for use in various parts of the cleanup process.

Because the gasification process takes place at elevated temperature, the vessel is generally cooled. Heat rejected there can be used to raise high-pressure steam for the steam cycle. Additional steam may be raised when cooling the hot fuel gas leaving the gasifier. This would be the case if a low-temperature cleanup process were to be used.

BASIC GASIFICATION COAL COMBINED CYCLE POWER SYSTEM

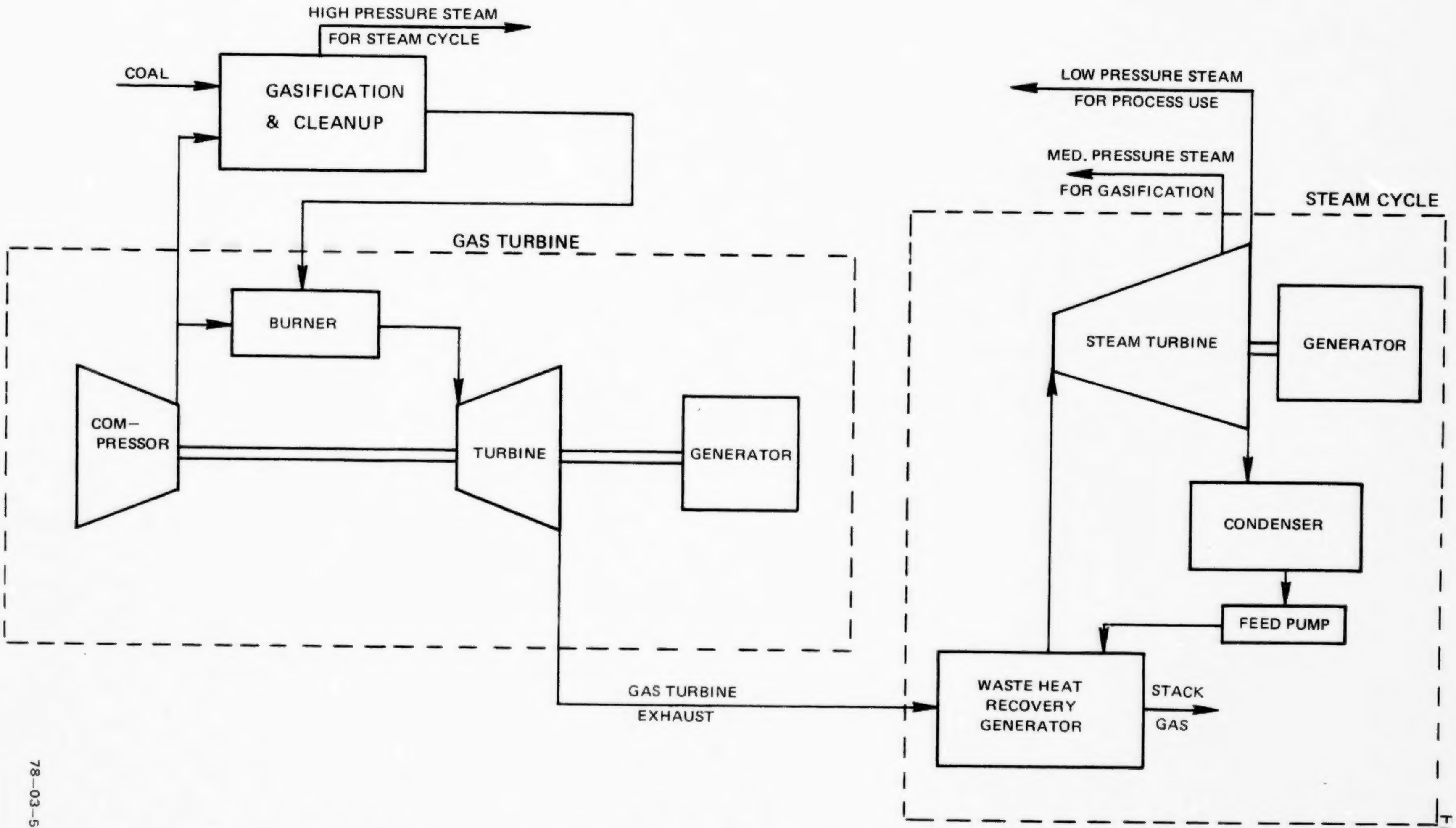

HIGH PRESSURE STEAM
FOR STEAM CYCLE
COAL
GASIFICATION & CLEANUP
GAS TURBINE
BURNER
COM–PRESSOR
TURBINE
GENERATOR
GAS TURBINE EXHAUST
LOW PRESSURE STEAM
FOR PROCESS USE
MED. PRESSURE STEAM
FOR GASIFICATION
STEAM CYCLE
STEAM TURBINE
GENERATOR
CONDENSER
FEED PUMP
WASTE HEAT RECOVERY GENERATOR
STACK
GAS

FIG. 5–1

78–03–54–1

The remainder of the plant is a conventional combined-cycle. Gas turbine exhaust is used to raise some steam for the steam turbine and to superheat that steam as well as that raised in the fuel processing system. The steam conditions are dependent on the available temperature levels and vary with gas turbine and fuel processing system parameters.

Gasification and Cleanup

The fuel processing part of the power plant is shown in simplified form in Figure 5-2. It assumes the use of a low-temperature cleanup system. In the coal handling section, coal is processed as required by the gasifier. It is fed to the gasifier at ambient condition. The gasifier must include lock-hoppers or other devices to feed the coal to the gasifier at pressure. Steam and air are fed to the gasifier as required and jacket heat is used to generate high-pressure steam for the steam cycle. Ash removed from the gasifier may require cooling or other handling facilities not shown here.

Gas leaving the gasifier must be cooled to permit removal of ammonia, hydrogen sulfide and particulates, at least in the case of low-temperature systems. In the cooling process heat may be used to raise steam and/or to reheat the clean fuel gas. This is discussed in Section 8 under thermal integration.

The water scrub consists of a particulate removal section and a section designed for ammonia removal. The particulate slurry produced, in most cases, can be recycled through the gasifier. The sour water, containing ammonia, carbon dioxide and hydrogen sulfide, is processed to remove ammonia. The remaining gases are sent to the sulfur recovery unit.

Sulfur in the fuel gas is removed in the H_2S absorber. The degree of sulfur removal and the amount of other gases absorbed differs among the various processes. These absorbed gases are stripped and sent to the sulfur recovery unit for recovery of elemental sulfur. While the degree of sulfur removal from the fuel gas is of prime importance, the amount of steam or power needed in the stripping operation and the concentration of H_2S in the feed to the sulfur recovery unit are important in measuring performance of the desulfurization process.

GASIFICATION

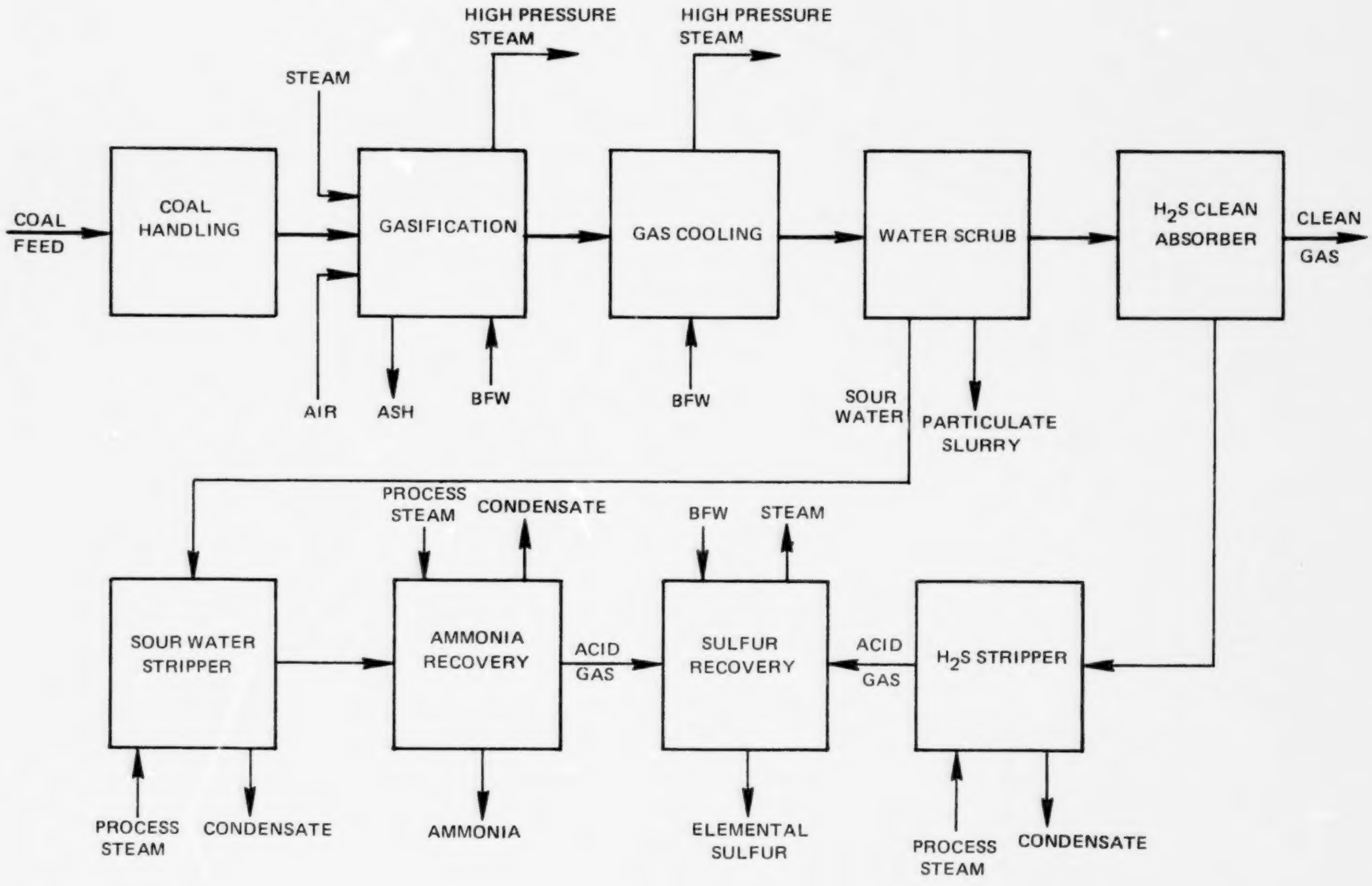
A high pressure, air-blown gasifier appears to be a logical choice for use with a combined cycle generating system. However, other types can be adapted and one goal of the current study is to compare the merits of air-vs. oxygen-blown operation.

Chemical Reactions

In the production of low-Btu gas, the highest yield would be achieved if all of the carbon were converted to carbon monoxide and all hydrogen were released in the molecular form. In practice many competing reactions exist and

FIG. 5-2

GENERAL FUEL PROCESSING SCHEMATIC


COAL FEED
COAL HANDLING
STEAM
AIR
GASIFICATION
ASH
BFW
HIGH PRESSURE STEAM
GAS COOLING
BFW
HIGH PRESSURE STEAM
WATER SCRUB
SOUR WATER
PARTICULATE SLURRY
H_2S CLEAN ABSORBER
CLEAN GAS
SOUR WATER STRIPPER
PROCESS STEAM
CONDENSATE
AMMONIA RECOVERY
PROCESS STEAM
CONDENSATE
AMMONIA
ACID GAS
SULFUR RECOVERY
BFW
STEAM
ELEMENTAL SULFUR
ACID GAS
H_2S STRIPPER
PROCESS STEAM
CONDENSATE

78-03-54-2

both CO_2 and H_2O are formed. The process is quite complex and can be viewed as a number of sequential reactions (Reference 5-1). First, some of the fuel is oxidized to form CO_2 and H_2O

$$C + O_2 \rightarrow CO_2 \tag{5-1}$$

$$2H_2 + O_2 \rightarrow 2H_2O \tag{5-2}$$

These reactions release a significant amount of heat which enables the highly endothermic gasification reactions to take place.

$$C + H_2O \rightarrow CO + H_2 \tag{5-3}$$

$$C + CO_2 \rightarrow 2CO \tag{5-4}$$

These are commonly known as the steam-carbon and Boudouard reactions. Boudouard equilibrium is generally used as a test for possible carbon formation. It determines the minimum CO_2 concentration.

In addition, methane can be formed by the reaction of hydrogen with carbon or carbon monoxide. Heating of the coal also results in devolatilization which produces carbon, methane and other hydrocarbons. At low temperatures, these hydrocarbons do not react and appear as heavy tars and oils in the gasifier product.

When these gases coexist at high temperature, equilibrium is generally determined by the water gas shift reaction,

$$H_2 + CO_2 \rightleftarrows H_2O + CO \tag{5-5}$$

which reaction is generally close to equilibrium.

Equilibrium concentration of methane can be calculated from the gas phase reaction:

$$CH_4 + H_2O \rightleftarrows CO + 3H_2 \tag{5-6}$$

However, measured concentrations are generally greater than would be predicted by equilibrium.

Formation of methane in the gasifier is of major importance when the objective of the gasification plant is to produce pipeline quality gas. When the production of fuel gas is the objective, the major consideration is the efficient conversion of the energy in the coal to chemical energy (heating value) of the product gas. This can be achieved with gas containing hydrogen and carbon monoxide as the major combustible components rather than methane.

Sulfur and Nitrogen Compounds--

Reactions involving sulfur and nitrogen are important from an environmental standpoint. Sulfur content of American coal varies widely. Western coals are generally low in sulfur while Eastern coals have relatively high sulfur content. Because of the reducing atmosphere in the gasifier, most of the sulfur is converted to hydrogen sulfide and carbonyl sulfide. The proportions of each are related by the reaction:

$$COS + H_2O \rightarrow CO_2 + H_2S \qquad (5\text{-}7)$$

At equilibrium, H_2S generally accounts for approximatly 95 percent of the total sulfur.

The conversion of coal nitrogen during gasification is more complex. Where oil and tar are produced, some of the nitrogen will appear in pyridine compounds. The remainder of the nitrogen will be converted to ammonia or molecular nitrogen. At high temperatures, some hydrogen cyanide may be formed.

Chemical Equilibrium--

Gasification processes vary greatly in the degree to which the reactions approach equilibrium. High-temperature and catalytic action within the gasifier result in a close approach to equilibrium. Sophisticated computational techniques are available to calculate the equilibrium concentration of virtually all potential products. The results of such a calculation are presented in Section 8 for the gasifiers studied under this contract.

Oxygen vs Air-Blown Gasification

The basic feed streams needed for gasification are coal, steam and a source of oxygen. Most processes can be designed to run with either pure oxygen or air as the source of oxygen. Processes used in producing pipeline gas are oxygen-blown and use steam to generate hydrogen for methanation. When oxygen is used, reactions are sufficiently exothermic as to make the addition of steam necessary to keep the temperature within practical limits . With air, the nitrogen diluent provides a quenching effect and complete gasification can theoretically be achieved with minimal steam used to reduce reaction temperature. The role of steam is critical in combined cycle power generation and is addressed in Section 8.

The choice between oxygen and air is usually determined by the end use of the gas. Oxygen is required when pipeline or high-Btu gas is desired. There, the presence of nitrogen would be unacceptable. Where transportation and interchangeability with natural gas are not a requirement, then medium- and low-Btu gas can be considered. Oxygen-blown operation will produce a medium-Btu gas with a heating value in the range of 300 Btu/SCF. An example of the difference in composition is given in Table 5-1 for a BCR-type gasifier. For air-blown operation, a value of 150 Btu/SCF is typical since the nitrogen diluent accounts for approximately 50 percent of the product gas. The primary fuel constituents of both are hydrogen and carbon monoxide which each have a heating value of approximately 325 Btu/SCF.

TABLE 5-1

COMPARISON OF AIR- VS OXYGEN-BLOWN GASIFICATION

BCR-Type Gasifier

Coal - Illinois No. 6 - 700,000 lb/hr

	AIR-BLOWN		OXYGEN-BLOWN	
	Mols/Hr	Mol %	Mols/Hr	Mol %
CH_4	3,775	3.37	4,522	6.04
H_2	15,315	13.68	22,098	29.50
CO	32,190	28.75	26,459	35.33
CO_2	3,396	3.03	9,369	12.51
H_2S	751	0.67	770	1.03
COS	76	0.07	77	0.10
N_2	53,753	48.02	336	0.45
NH_3	479	0.43	479	0.64
H_2O	2,213	1.98	10,787	14.40
	111,948		74,897	

	AIR-BLOWN	OXYGEN-BLOWN
HHV-Btu/SCF	171.2	270.8
Oxidant/Coal Ratio	2.78	0.594
Steam/Coal Ratio	0.144	0.597
Transport Gas/Coal Ratio	0.088	0.088
Cold Gas Efficiency	83%	85.7%

When used as a fuel, the primary difference between low- and medium-Btu gas is the flame temperature. Because of the nitrogen diluent, the stoichiometric flame temperature of the low-Btu gas can be as much as 800 F lower than that of the medium Btu gas. Of interest is the fact that the stoichiometric flame temperature of methane or high-Btu gas is lower than that of medium Btu gas. Methane requires four times as much air as does an equivalent volume of CO or H_2 (the major constituents of medium-Btu gas) while having a heating value of just over three times that of those fuels.

Flame temperature can be very important when retrofitting existing industrial processes. For that purpose medium-Btu gas would be selected. Also, it has the advantage of being more economically transported. However, in the gas turbine, flame temperature is not important, in fact the high flame temperature associated with medium-Btu gas may result in a NO_x problem discussed in Section 9, and selection can be based on other factors.

Air-blown operation is generally favored because oxidant for the gasification process can be bled from the gas turbine compressor. The nitrogen is returned to the gas turbine with the fuel gas and produces useful work as it is expanded through the turbine. Oxygen-blown operation requires a separate oxygen compressor in addition to an air separation plant. It does, however, result in smaller cleanup equipment size since the fuel gas volume is approximately half that of the air-blown system. This, plus the possible synergism of industrial utilization of the fuel gas, could make the oxygen-blown gasifier a viable alternative.

Types of Gasifiers

While the reactions are common to all gasifiers, i.e., partial combustion followed by reducing type reactions, the means of achieving them differ widely. Gasifiers can be classified in four main categories:

(a) Moving Bed
(b) Fluidized Bed
(c) Entrained Bed
(d) Molten Bath

Within each category, gasifier characteristics can vary due to operating conditions (temperature and pressure), method of coal feeding, method of ash removal and other important aspects. However, it is a convenient categorization and represents a primary design difference between the processes.

Moving Bed Gasifier--

This category is typified by the Lurgi gasifier although, in this report, data for the Bureau of Mines stirred bed gasifier have been used. It is sometimes referred to as a "Fixed Bed" as it involves a bed of coal supported by a grate. Oxidant and steam are introduced below the grate and move rather slowly up through the bed. Product gas is removed from the top while ash is discharged at the bottom.

Combustion temperature in "dry bottom" operation is controlled to below the ash fusion point. Temperature of the gas leaving the top of the unit is approximately 1000 F. Because of the relatively low temperatures, the effluent gas contains significant amounts of condensible oil and tar vapors. As a result, the gas is usually quenched with water to condense and remove these constituents. This is a major disadvantage in combined-cycle operation since most of the sensible heat in the off-gas is lost.

The process is sensitive to type and size of the coal feed. Caking coals tend to form clinkers and a stirring device is provided to break up any lumps that may form. Fines are undesirable and must be briquetted or used elsewhere. Coal is fed by lock-hopper. The high coal inventory in the gasifier provides stability against variations in coal flow and composition.

A recent advance in moving bed gasifiers, which is now in the development stage, is the "slagging" mode of operation. In this operation, the gasifier is oxygen-blown and the combustion zone is controlled at a temperature which is above the melting point of the ash. Ash is removed from the gasifier in molten form. This type of operation reduces the steam requirement and increases the gasification rate relative to "dry bottom" gasifiers.

Fluidized Bed Gasifier--

As the name implies, solids in the gasifier are suspended in a fluid state by an upflow of gas through the gasifier. As is characteristic of a fluidized bed, temperature is quite uniform and there are no hot spots. Thus, temperatures can be maintained below the ash fusion point while being high enough to minimize tars and oils in the off-gas (generally 1600-1800 F). However, this is a problem in many fluid bed concepts. Also, since there must be a considerable carbon inventory, ash disposal is a problem. The fluid bed is well mixed and ash removal is a matter of extracting a part of the well-mixed bed. As a result, the ash removal stream contains a substantial amount of unreacted carbon.

A problem common to gasifiers of this type is the inability to use strongly caking coal without pretreatment. However, they are tolerant to variations in coal size and require less coal processing equipment.

Many of the problems associated with fluidized bed gasification are claimed to be eliminated in the ash-agglomerating gasifier under development at IGT. Relatively low amounts of carbon are contained in the ash and no tar is expected in the off-gas. Operation of that gasifier is described in Section 6.

Entrained Flow Gasifiers--

Gasification by means of an entrained flow process involves suspension of relatively small coal or coal char particles in a high velocity gaseous medium Residence time of the coal particles in the gasifier is relatively low and high temperatures are used to maximize gasification rate. As a result, entrained flow gasifiers operate under ash slagging conditions.

The atmospheric pressure Koppers-Totzek (K-T) gasifier is probably the best working example of a single-stage entrained flow process. It is oxygen-blown and reaction temperature is in excess of 2600 F. The high temperature results in high reaction rates and the off-gas is reported to be very close to equilibrium. The coal feed is pulverized and the gasifier is not affected by the use of caking coals. Application of the K-T process to combined cycle power generation is discussed in Section 6.

In a two-stage entrained flow gasifier, coal is injected into an upper stage where it is partially gasified by hot gases produced in the lower stage. Gas and char leave the upper stage at a temperature near 1800 F. Char is separated from the gas and injected into the lower stage together with steam and an oxygen containing gas. The char is partially combusted at temperatures of 2800 to 3000 F. Molten ash slag is separated from the gases formed and is removed from the bottom of the lower stage.

As with the single-stage gasifier, any grade of coal can be used since the pulverized particles are in the reaction zone. The carbon content in the ash is quite low because of the high reaction temperature. The capacity of entrained flow gasifiers is much higher than that of moving or fluidized bed gasifiers since the flow is not restricted by bed characteristics.

Molten Bath Gasifiers--

In this type of process, the reactants are contacted in a high temperature molten fluid. The melt serves to disperse the reactants and provides heat transfer and thermal storage for the process. It can also have a catalytic effect on gasifier reactions and/or combine chemically to remove sulfur from the product gas. As in the entrained flow process, the gasifier can handle all types of coal. It has the additional advantage of being relatively insensitive to coal size, thus reducing the processing requirements. Ash is withdrawn with the melt (the ash forms the molten bath in some cases) and contains on the order of 1 percent unreacted carbon.

Gas leaving the process is free of tars but may contain vapors from the melt. In the case of a molten salt type gasifier, the vapors could be harmful and provision must be made to insure their removal.

CLEANUP SYSTEMS

Fuel gas cleanup systems consist of processes for sulfur removal, nitrogen compound (ammonia) removal, and particulate removal. Sulfur removal processes are generally divided into two categories. These are:

1) Low-temperature desulfurization
2) High-temperature desulfurization

Low-temperature desulfurization processes require cooling of the dirty gas and operate at temperatures below 250 F. At these temperatures, it is possible to use water as a means of particulate and ammonia removal. High-temperature desulfurization processes operate at or near gasifier exit conditions and their use will require other means of nitrogen and particulate control. Control of nitrogen and particulates in systems with high-temperature desulfurization is discussed in Section 8.

Low-Temperature Desulfurization Processes

Low-temperature processes for desulfurizing raw producer gas are commercially available and have been widely used for natural gas sweetening and treating synthesis gas in the chemical process industry; e.g., the manufacture of ammonia, methanol and oxo-alcohols. These systems normally operate below 250 F and are commonly classified into the following categories:

Chemical Solvent Processes
Physical Solvent
Direct Conversion Processes
Dry Bed Processes

A literature survey resulted in a list of 38 low-temperature processes. A summary of the important characteristics of some representative processes is given in Table 5-2. Figure 5-3 illustrates the schematic for a tyical absorption - stripping process for low-temperature acid gas removal.

Chemical Solvent Processes--

Chemical solvent processes employ aqueous solutions of organic and/or inorganic agents which are capable of forming "complexes" with the acid gas components, notably H_2S and CO_2, present in the raw gas stream. The absorption solution is regenerated by decomposing the "complex" at elevated temperature thereby releasing the acid gases for subsequent recovery. The solution is recycled for further absorption. These processes are essentially insensitive to the partial pressure of acid gases in the feed and generally exhibit little or no selective absorption of H_2S over carbon dioxide. The chemical processes may be sub-divided into those using amine scrubbing solutions and those based on alkali scrubbing solutions.

The principal reactions involved in gas sweetening with amine solutions (10-30 percent weight) may be represented as:

$$RNH_2 + H_2S \rightarrow RNH_3\,HS \qquad (5\text{-}8)$$

$$RNH_2 + CO_2 + H_2O \rightarrow RNH_3\,HCO_3 \qquad (5\text{-}9)$$

Monoethanolamine (MEA) will easily reduce the H_2S content below 4 ppm; however, it is not considered selective, even though the rate for CO_2 absorption is less than for H_2S. The principal disadvantage of MEA is that it will react with COS and CS_2 forming nonregenerable compounds. Diethanolamine (DEA) will not react with these contaminants and is favored for service where

TABLE 5-2

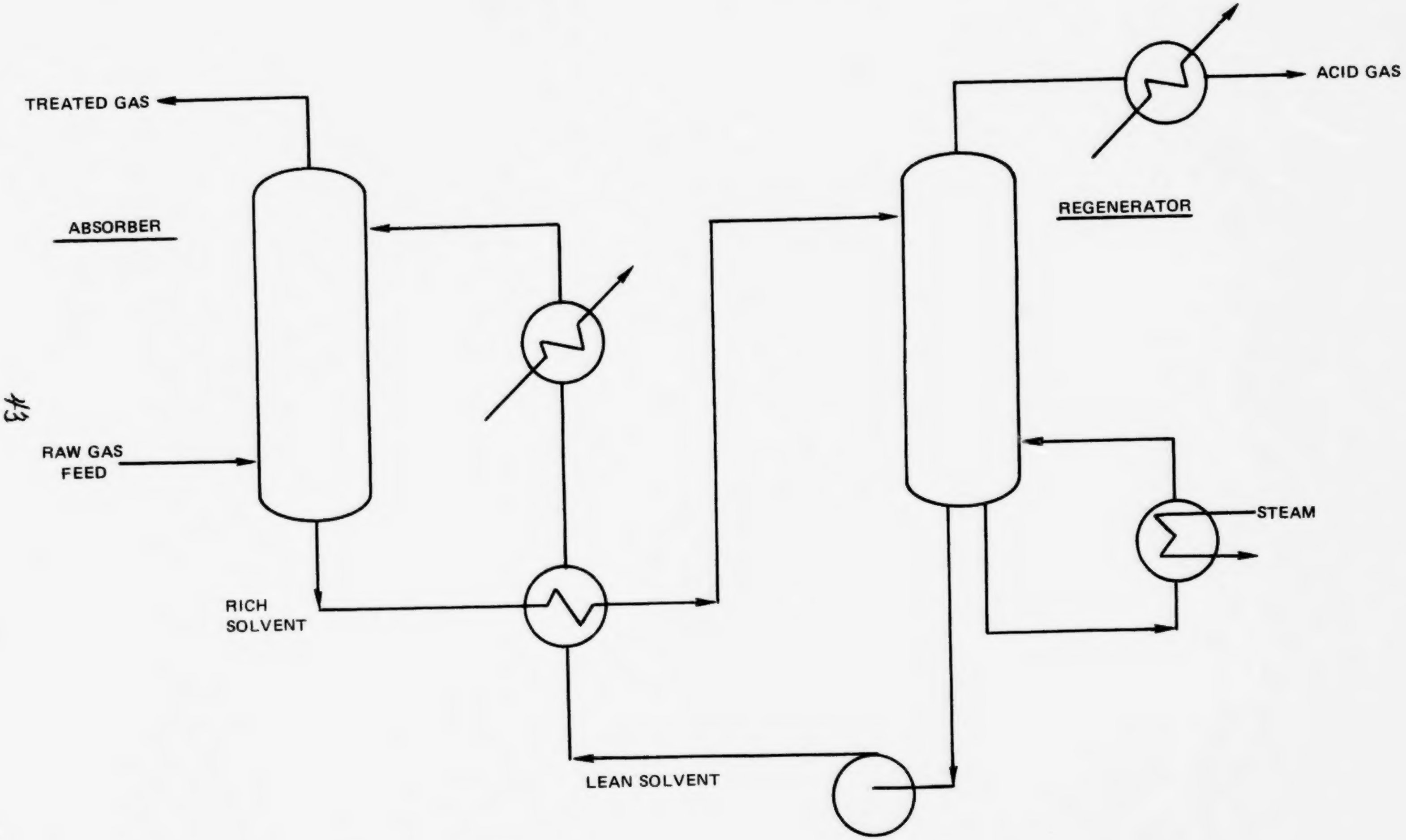
LOW TEMPERATURE CLEANUP PROCESSES

Basis: 8400 tons/day Illinois No. 6 Coal Fed to BCR Gasifier, or 6700 ppm of Influent H_2S

Process	Absorbent	Type of Absorbent	Temp. °F	Pressure	Efficiency of S Removal		Absorbent Characteristics				Form of Sulfur Recovery	Status
					% H_2S Influent	Effluent H_2S ppm	Life	Regeneration	Selectivity toward	Make up rate		
Chemical solvent type												
1. MEA	Monoethanolamine	Aqueous solution	80 to 120	Insensitive to variation in pressure	99	~100		Thermal	Forms non-regen. comp. with COS, CS_2	50 to 100%	As H_2S gas	Commercial
2. DEA	Diethanol amine	Aqueous solution	100 to 130	Insensitive to variation in pressure	99	~100		Thermal	Absorbs CO_2, does not absorb COS, CS_2	< 5%	As H_2S gas	Commercial
3. TEA	Triethanolamine	Aqueous solution	100 to 150	Insensitive to variation in pressure	99	~100		Thermal	H_2S	< 5%	As H_2S gas	Commercial
4. Alkazid	Potassium dimethyl amino acetate	Aqueous solution	70 to 120	Insensitive to variation in pressure 1 - 80 atm	99	~100		With steam	H_2S		As H_2S gas	Commercial
5. Benfield	Activated potassium carbonate solution	Aqueous solution	150 to 250		99	H_2S + COS ~100	Unlimited No degradation	With steam	H_2S is high		As H_2S gas	Commercial
6. Catacarb	Activated potassium carbonate solution	Aqueous solution	150 to 250	Insensitive to variation in pressure generally > 300 psi	99	H_2S + COS ~100		With steam	H_2S - partial also absorbs COS, CS_2	< 5%	As H_2S gas	Commercial
Physical solvent type												
7. Sulfinol	Sulfolane + Diisopropanoamine	Organic solvent	80 to 120	High pressure preferred	99	H_2S + COS ~100		Low pressure heating or with steam	H_2S, and also absorbs COS, CS_2 and mercaptans		As H_2S gas	Commercial
8. Selexol	Polyethylene glycol ether	Organic solvent	20 to 80		99	H_2S + COS ~100			H_2S also absorbs COS		As H_2S gas	Commercial
9. Rectisol	Methanol	Organic solvent	< 0		99	~100			H_2S			Commercial
Direct conversion												
10. Stretford	Na_2CO_3 + anthraquinone sulfonic acid	Alkaline solution			99.9	~10			H_2S	50 to 100%	Elemental sulfur	Commercial
11. Townsend	Triethylene glycol	Aqueous solution	150 to 250		99.9	~10			H_2S		Elemental sulfur	
Drybed type												
12. Iron sponge	Hydrated Fe_2O_3	Fixed bed	70 to 100		99	H_2S + COS ~100			H_2S and also towards COS, CS_2 and mercaptans		Elemental sulfur	Commercial

FIG.5–3

TYPICAL LOW – TEMPERATURE ACID GAS REMOVAL UNIT


TREATED GAS
ABSORBER
RAW GAS FEED
RICH SOLVENT
LEAN SOLVENT
REGENERATOR
ACID GAS
STEAM

R1–19–2

COS and CS_2 are likely to be present. Like MEA, DEA solutions are not selective for H_2S and will seldom reduce the H_2S content below 100 ppm. Tertiary amines, such as triethanolamine and methyl-diethanolamine, while not as reactive as the other amines, have the advantage of being selective towards H_2S removal. The tertiary amines are two to four times more costly and find little application in industrial gas sweetening.

The alkali scrubbing system may be represented by the following chemical reactions:

$$M_2CO_3 + H_2S \rightarrow MHS + MHCO_3 \qquad (5\text{-}10)$$

$$M_2CO_3 + CO_2 + H_2O \rightarrow 2\ MHCO_3 \qquad (5\text{-}11)$$

A number of processes have been based on these reactions, the Benfield and Catacarb representing the most advanced versions.

The earlier processes, such as Seaboard and Vacuum Carbonate, were based on dilute solutions of sodium carbonate (3-4 percent weight) and were capable of removing 80-90 percent of the H_2S. Regeneration in the Seaboard process was by air resulting in a dilute acid gas stream while the Vacuum Carbonate system used vacuum regeneration with steam. These processes were superseded by the hot potassium carbonate system. In the "hot pot" processes, an aqueous solution of 25-35 weight percent $K_2\ CO_3$ is used to absorb acid gases at temperatures in the range of 200-250 F. With low H_2S/CO_2 ratios, the process is capable of sweetening the gas to 5 ppm. Some selective H_2S absorption can be achieved by taking advantage of the relatively slow rate for CO_2 absorption. In addition to removing H_2S and CO_2, the process can remove COS and CS_2 by hydrolysis of these components to CO_2 and H_2S. The Catacarb and Benfield processes are improved versions of the Bureau of Mines "hot pot" systems. They employ activators to increase the rate of absorption thereby decreasing the required solution circulation rate. Disadvantages of the hot potassium carbonate systems are the relatively high steam consumption for regeneration, sensitivity to operating pressure and possible inability to handle mercaptans and thiophene.

Not listed in Table 5-2, the tripotassium phosphate process was developed by the Shell Oil Company specifically for H_2S removal via the reaction:

$$K_3PO_4 + H_2S \rightarrow K_2HPO_4 + KHS \qquad (5\text{-}12)$$

The nonvolatility of the agent, its nonreactivity with COS and CS_2, and partial selectivity toward H_2S in the presence of CO_2 gives it certain advantages over the amine systems. However, when operated for high H_2S selectivity, the process only gives about 90 percent removal efficiencies. Conversely, with high H_2S removal the steam consumption becomes excessive due to CO_2 absorption.

Physical Solvent Processes--

Physical solvent processes all use organic solvents to remove acid gases. The physical absorption is directly proportional to the partial pressure of the acid gas components. These processes are most applicable to high-pressure gas treating where appreciable quantities of sour gases are present. After absorption, the "loaded" solvent is regenerated by heat and/or pressure reduction giving a concentrated stream of H_2S plus CO_2 and a recyclable lean solvent. Due to the higher solubility of H_2S in these organic solvents, selective absorption of H_2S over CO_2 can be achieved. In general, these processes have two major disadvantages: the solvents have a great affinity for absorbing heavy hydrocarbons (C_5+) which contaminate the gas stream fed to sulfur recovery units; and the solvents are quite expensive so that large solvent losses cannot be tolerated.

As a group, these processes were developed for bulk removal of acid gases but, for low H_2S concentration, they are capable of giving a sweetened gas having less than 5 ppm H_2S. In order to maximize the solubility of acid gases and minimize solvent loss through vaporization, the processes are generally operated at or below ambient temperature. In addition to removing H_2S and CO_2, the solvent processes are all capable of removing COS, CS_2 and mercaptans without solvent degradation. They also dehydrate the gas to a low dew point. The low heats of solution for acid gases result in appreciably lower steam requirements for solvent regeneration compared to steam requirements for the chemical solvents.

The Sulfinol process is unique in that it combines the characteristics of a solvent process and an amine process. The physical absorbent, Sulfolane, gives high acid gas loadings at high acid gas partial pressures, giving it bulk removal capacity and the chemical absorbent, DIPA,* reduces residual acid gases to very low values. However, the presence of the chemical solvent reduces the H_2S selective H_2S absorption capability for this system compared to the straight solvent processes.

Direct Conversion Processes--

Two types of processes fall into this category:

a. those based on oxidation reduction reactions, and

b. those based on the stoichiometric reaction of H_2S with SO_2 in the presence of a solvent.

The first type involves the absorption of H_2S in alkaline solutions containing oxygen carriers. The H_2S is subsequently oxidized to elemental sulfur by air fed to the regenerator. There the sulfur product is flotated and collected as a froth at the regenerated solution interface. Processes of this type are in common use in Europe for removal of H_2S and sulfur recovery from manufactured gases and coke-oven gas. The Ferrox and Manchester processes employ a suspension of iron oxide in an aqueous solution of sodium carbonate to absorb H_2S. With multistage absorption, essentially complete removal of H_2S is obtainable; however, the product is of low quality due to salt contamination. Chemical replacement costs are high. Both the Thylox and Giammarco Vetrocoke (References 5-2, 5-3) processes use alkaline solutions of arsenates and are capable of reducing the H_2S to less than 1 ppm. Partial removal of COS, CS_2, and mercaptans is also possible. Again the sulfur product is contaminated and the use of arsenates makes these processes potentially hazardous The Stretford and Takahax processes are similar in that alkaline solutions of quinone sulfonic acids are employed. The addition of vanadium salts increase the rate of oxidation of hydrosulfide to sulfur resulting in higher solution loadings. Close to 100 percent H_2S removal is claimed for these processes as is a high purity (99%) sulfur product. However, substantial amounts of thiosulfates are formed resulting in sludge deposition and corresponding chemical makeup.

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* di-isopropanolamine

Generally speaking, the low solution loadings exhibited by this group of processes make them uneconomical for treating large volumes or very sour gas streams. They are best suited for sour gases containing less than 1.0 percent H_2S with sulfur production under 20 tons/day. These processes, as well as those in the following group, are almost totally selective for H_2S removal.

In the second group of direct conversion processes are those in which H_2 is absorbed in a solvent and converted to elemental sulfur by the Claus type reaction with SO_2.

$$2\ H_2S + SO_2 \rightarrow 3\ S + H_2O \tag{5-13}$$

The Townsend process uses an aqueous solution of an organic solvent, such as triethylene glycol, to sweeten the gas, dehydrate the gas and convert H_2S to elemental sulfur. A portion of the product sulfur is burned to SO_2 which is absorbed by fresh solvent and the SO_2 rich solvent is used to contact the sour gas. The IFP, Nalco, and Deal processes operate in a similar manner employing other solvents. While a high purity (99.7%) sulfur product is claimed, none of these processes have been commercialized.

Dry Bed Processes--

These sweetening processes are based on absorption of acid gases by a fixed bed of solid absorbent. Due to their low absorbent loading, they are applicable to gases containing low concentration of H_2S and mercaptans, perhaps less than 500 ppm. These processes can be subdivided into the iron oxide processes and the various molecular sieve processes.

The iron oxide or dry box process is one of the oldest processes known for removing sulfur compounds from industrial gases. In the iron sponge systems, wood shavings impregnated with hydrated ferric oxide are used to absorb H_2S:

$$2\ Fe_2O_3 + 6\ H_2S \rightarrow 2\ Fe_2S_3 + 6\ H_2O \tag{5-14}$$

Regeneration of the absorbent is carried out with air:

$$2\ Fe_2S_3 + 3O_2 \rightarrow 2\ Fe_2O_3 + 6S \qquad (5\text{-}15)$$

This process is best suited for small to medium gas volumes with low sulfur contents; otherwise the sponge bed life would be too short to be economical. The process is selective towards H_2S and mercaptans and will partially remove COS and CS_2. Sweetened gas of less than 5 ppm H_2S is easily obtained However, sulfur recovery would not be economical when using the iron sponge system.

Molecular sieves can be made to have pore sizes which will permit selective absorption of H_2S over CO_2. These processes are characterized by the various regeneration schemes employed; i.e., via hot combustion gases, hot SO_2 gas as in the Haines process, or hot air. In the latter two modes, elemental sulfur is produced via the oxidation of the absorbed H_2S. The sieve processes also appear to be economically attractive for small to medium gas volumes having low H_2S content. Additionally, for efficient H_2S removal, the raw sour gas should have a water content below 20 lb/MMSCF since water will also be absorbed by the molecular sieve structure.

Selection Considerations--

Low-temperature desulfurization systems for application in low-Btu fuel gas plants will have to treat large volumes of sour gas, 500-1000 MMSCFD, having total sulfur content in the range of 0.4 to 1.0 percent by weight. In addition to H_2S, the raw gas will contain CO_2, COS, CS_2, probably mercaptans, cyanides and heavy hydrocarbons. Of the types of processes described above, it is evident that the liquid scrubbing processes using physical solvents and some chemical solvents are the best suited. These processes are currently available and can easily reduce the sulfur content of the gas to 100 ppm. Such a gas, when combusted, would result in SO_2 emissions well below present EPA regulations for conventional steam stations. As such, these processes are capable for serving both first-generation and second-generation coal gasification plants and should meet any future regulations for SO_2 emissions.

High Temperature Cleanup Systems

High-temperature systems for sulfur removal are not presently available in commercial scale although there are several in various stages of development. Active work involves use of limestone and dolomites which have potential in the range of 1500-2000 F. Other systems receiving attention employ iron oxides, molten salts, and liquid metals. These systems operate by chemical reaction of the absorbent with sulfur compounds in the gas, forming the corresponding metal sulfides. The degree of desulfurization attainable depends in part on the chemical equilibria for the particular system at the operating conditions. As with low-temperature processes, economics dictate that the sulfided absorbent be regenerated for reuse.

The only commercial experience with high-temperature desulfurization reported in the literature is that of the Frodingham Desulfurization Process (Refs. 5-4, 5-5). This process employed fluidized beds of pulverized iron oxide operating at 650 F to 800 F. In the early 1960's, a commercial plant treating 32 MMCFD of crude coke oven gas containing 1.0 percent H_2S was operated at the Exeter Works of the South Western Gas Board. Essentially complete (99.9%) removal of H_2S was achieved with 90 percent removal of all organic sulfur compounds other than thiophene. The sulfided absorbent was regenerated with air at 1000-1100 F resulting in a SO_2 stream which was subsequently converted to sulfuric acid. Major difficulties were experienced in the solids handling system which produced fine oxide dust resulting in excessive losses of the absorbent. In addition, operation of the sulfuric acid plant was erratic due to low SO_2 concentrations in the regenerator off-gas.

Several processes currently under development which may prove commercially viable for use with second generation gasification systems are listed in Table 5-3.

Bureau of Mines Sintered Iron Oxide Process-- (References 5-6, 5-7, 5-8, 5-9.)

This process, under development at the Morgantown Energy Research Center, is based on a sintered absorbent consisting of a mixture of iron oxide (Fe_2O_3) and fly ash. This sorbent satisfied the primary requirements for high-temperature sulfur removal in that it is readily available and inexpensive, it has reasonable absorption capacity for sulfur, can be regenerated for repeated use and is resistant to fusion and disintegration over the operating temperature range of 1000-1500 F. The absorbent is prepared by mixing iron oxide with "as received" fly ash to a total iron oxide content of about 35 percent. Iron oxide contents above 40 percent were unsatisfactory because the fusion temperature was lowered within the operating range. The mixture is extruded into 1/4" x 3/8" pellets and then sintered to develop the required hardness.

Reaction chemistry is reported in Reference 5-7 as follows. During absorption, two iron sulfides are produced, FeS and FeS_2, with the empirical composition approaching $FeS_{1.5}$:

$$Fe_2O_3 + 3H_2S \rightarrow 2Fe_{1.5} + 3H_2O \qquad (5\text{-}16)$$

During regeneration, air oxidizes the FeS/FeS_2 to Fe_2O_3 and SO_2.

$$6FeS_{1.5} + 13O_2 \rightarrow 2Fe_3O_4 + 9SO_2 \qquad (5\text{-}17)$$

$$2Fe_3O_4 + 1/2\ O_2 \rightarrow 3Fe_2O_3 \qquad (5\text{-}18)$$

Excess oxygen in the air from the reaction that produces the Fe_3O_4 converts it to Fe_2O_3.

TABLE 5-3

HIGH TEMPERATURE CLEANUP PROCESSES

Basis: 8400 tons/day Illinois No. 6 Coal Fed to BCR Gasifier, or 6700 ppm of Influent H_2S

Process	Absorbent	Type of Bed	Temp. °F	Pressure	Efficiency of S Removal		Absorbent Characteristics				Form of Sulfur Recovery	Energy Required		Status
					%H_2S Influent	Effluent H_2S ppm	Life	Regeneration	Selectivity toward	Make up rate		Elec. kw	Other stu	
1. Bureau of Mines	Sintered pellets of Fe_2O_3 (25%) and fly ash	Fixed bed	1000 to 1500	Insensitive to variation in pressure	~95	~350	>174 cycles Wt loss < 5%	With air	H_2S, COS	<5%	As SO_2 gas			Pilot
2. Babcock and Wilcox	Fe_2O_3	Fixed bed	800 to 1200	Insensitive to variation in pressure	~99	~75					As 12-15% SO_2 gas			Experimental
3. CONOCO	Half calcined dolomite	Fluidized bed	1500 to 1800	~200 psia H_2S removal is high at low pressure	~95	~350		10-13% with steam and CO_2	H_2S, COS	1% of circulation rate	As H_2S gas to Claus process	96.360		Pilot
4. Air products	Calcined dolomite	Fixed bed	1600 to 2000	Insensitive to variation in pressure			minimum 5-6 cycles	80-90% with steam and CO_2	H_2S, COS		As H_2S gas to Claus process			Abandoned
5. Battelle Northwest	Molten carbonates (15% $CaCO_3$)	Solution	1100 to 1700	Atmospheric H_2S removal is high at low pressure, 5-6 psig	~95	~350		With steam and CO_2	H_2S, COS, fly ash		As H_2S gas to Claus process			Pilot
6. IGT - Meissner	Molten metal (proprietary)	Splashing contact	900		~98	~150		Electrolytic	H_2S, COS			9830		Conceptual

A practical problem during regeneration is the burnup of accumulated carbon deposits within the bed. This produces locally high temperatures that tend to harm the absorbent. Recent reports indicate that the fly ash can be replaced by silica to produce a more rugged material.

Concern over the possible absorption of hydrogen or the conversion of hydrogen and carbon monoxide to water and carbon dioxide by reaction with the ferric oxide is expressed in Reference 5-10. Equilibrium calculations indicate the potential for high sulfur removal levels, especially at low steam feed rates. Carbonyl sulfide absorption also appears to be practical.

An area that has not been investigated is the potential effect of the iron oxide bed on ammonia in the fuel gas. It is possible that it will catalyze the decomposition of ammonia which is generally assumed to be present in concentrations well in excess of those at equilibrium. This could be a very attractive feature since the presence of ammonia and consequent NO_x production is a major drawback in high-temperature desulfurization.

Conoco Half Calcined Dolomite Process--

This process (Reference 5-11) evolved as an adaption of the CO_2 Acceptor Process (Reference 5-12) for producing low-Btu fuel gas from coal and incorporates the use of a half-calcined dolomite acceptor for sulfur capture as studied by Squires and coworkers (References 5-13 to 5-16). Basically the process chemistry involves the following reaction:

$$[CaCO_3 \cdot MgO] + HS_2 \rightarrow [CaS \cdot MgO] + HO_2 + CO_2 \qquad (5\text{-}19)$$

A maximum operating temperature for this process is imposed by the partial pressure of carbon dioxide in the gas phase; i.e., the temperature should not exceed that at which the CO_2 partial pressure is equal to the decomposition pressure for $CaCO_3$ via the following endothermic reaction:

$$CaCO_3 \rightarrow CaO + CO_2 \qquad (5\text{-}20)$$

Should that temperature be exceeded, CO_2 would be released into the clean fuel gas and the regeneration of the unreacted CaO will make the process more complex. The effect of this temperature limitation on desulfurization performance is discussed in Section 8 where the various factors are quantified.

Although no data have been reported for COS adsorption by half-calcined dolomite, high COS removal efficiencies are predicted thermodynamically according to the reaction:

$$[CaCO_3 \cdot MgO] + COS \rightarrow [CaS \cdot MgO] + 2\,CO_2 \qquad (5\text{-}21)$$

The process as described by Conoco involves desulfurizing the raw gas in a fluidized bed of half-calcined dolomite acceptor at 1600-1700 F, according to reaction (5-19). The sulfided acceptor is regenerated by the addition of steam and CO_2 at reduced temperature, thereby reversing the absorption reaction. Regeneration is conducted in a fluidized bed at around 1300 F giving a dilute H_2S off-gas,.less than 10 percent (volume). Because the low H_2S off-gas content prohibits the direct use of a vapor-phase Claus unit for sulfur recovery, Conoco is proposing the use of a liquid phase sulfur recovery system based on the Wachenroeder reaction

$$2H_2S + H_2SO_3 \rightarrow 3S + 3H_2O \qquad (5\text{-}22)$$

All important features of the process have been confirmed (Ref. 5-11) in bench-scale studies. There remains to be determined the effect of scale-up as well as the effect of ammonia and the ability to remove particulates and alkali fumes from the effluent. The process is described in detail in Section 6 and the effect of gasifier operating conditions on sulfur removal examined in Section 8.

Air Products Fully Calcined Dolomite Process--

As with the Conoco process, this system employs dolomite as the sulfur acceptor. However, the acceptor is in the fully calcined form; i.e., $[CaO \cdot MgO]$, and the process therefore consists of three steps: absorption, regeneration and calcination.

Absorption of hydrogen sulfide takes place at around 1600-1700 F via the reaction with calcined dolomite:

$$[CaO \cdot. MgO] + H_2 \rightarrow [CaS \cdot MgO] \qquad (5\text{-}24)$$

For this system, the H_2S removal efficiency is independent of pressure but, for a given gas composition, it decreases with increasing temperature. From a practical viewpoint, there is a minimum temperature at which this process should operate; that being the temperature at which the decomposition pressure of $CaCO_3$ equals the partial pressure of CO_2 in the gas. Below this temperature carbon dioxide will also be absorbed according to the reaction.

$$CaO + CO_2 \rightarrow CaCO_3$$

While the carbonate can also react with H_2 in accordance with reaction (5-19), this is detrimental for two reasons:

a. Additional acceptor calciner capacity is required with an associated increase in heat input, and

b. The large exothermic heat of CO_2 absorption, 75,000 Btu/mol, will necessitate some means for heat removal from the absorption bed.

Like the Bureau of Mines and Conoco processes, COS removal appears to be thermodynamically attractive. Residual COS content may be estimated from the chemical equilibrium associated with the absorption reaction:

$$[CaO \cdot MgO] + COS \rightarrow [CaS \cdot MgO] + CO_2 \qquad (5\text{-}25)$$

Regeneration of the sulfided acceptor is conducted similarly to the Conoco process. Steam and carbon dioxide are reacted with the sulfided dolomite at 1100-1200 F resulting in a dilute H_2S off-gas and half-calcined dolomite.

$$[CaS \cdot MgO] + H_2O + CO_2 \rightarrow [CaCO_3 \cdot MgO] + H_2S \qquad (5\text{-}26)$$

Hydrogen sulfide in the regenerated off-gas may be converted to elemental sulfur via the liquid-phase Wachenroeder process or first concentrated and then fed to vapor-phase Claus units.

The regenerated half-calcined dolomite must be calcined before recycle to the subsequent absorption cycle. Calcination is effected at 1900 F with air to drive off carbon dioxide:

$$[CaCO_3 \cdot MgO] \rightarrow [CaO \cdot MgO] + CO_2 \qquad (5\text{-}27)$$

This reaction is endothermic and requires the use of fuel to preheat the air. Excessive temperature, above 2000 F, during calcination can result in deactivation of the acceptor.

There is currently no development activity in progress.

IGT-Meissner Process--

This process is being developed by the Institute of Gas Technology in conjunction with its U-Gas Process. The process, still in the conceptual stage, utilizes a splashing molten metal-gas contact to remove H_2S from the gas. The contact results in the formation of a metal sulfide which is then decomposed electrolytically to release H_2S and regenerate the molten metal for recycle. The operating temperature is 900 F and a high sulfur removal efficiency (98 percent) is projected. The molten metal absorbent is proprietary. The estimated characteristics given in Table 5-3 are preliminary (Reference 5-17). Further development is being directed toward establishing mass transfer rates.

Battelle Northwest Process--

The Battelle process (Reference 5-18) utilizes calcium carbonate, $CaCO_3$, dissolved in a tertiary mixture of alkali metal carbonates to remove H_2S at high temperature. The tertiary carbonate system, consisting of potassium carbonate, lithium carbonate, and sodium carbonate, has a eutectic melting point around 750 F. Under operating conditions it contains about 15 mol percent $CaCO_3$. Besides removing sulfur compounds from the gas stream, this solvent will also scrub out the fly ash constituents from the raw gas.

The system under study at Battelle contacts the molten salt and raw gas in a co-current flow venturi scrubber at temperatures from 1100 to 1700 F. Hydrogen sulfide is removed from the gas by chemical reaction with the carbonate solvent:

$$CaCO_3 + H_2S \rightarrow CaS + CO_2 + H_2O \qquad (5\text{-}28)$$

Unfortunately, due to the chemical complexity of the molten salt system, the equilibrium constant cannot be accurately predicted. Observed K values have been a factor of ten below the calculated values. Qualitatively, the H_2S removal efficiency improves with temperature and is inversely proportional to pressure. Experimental data at atmospheric pressure have indicated high H_2S removal, >94 percent, at salt loadings under 50 percent of capacity. Regeneration of the salt is conducted with steam and CO_2 at 1000-1100 F giving an off-gas having H_2S concentrations suitable as feed to a Claus unit for recovery of elemental sulfur.

This process is presently in the pilot plant stage where it will be demonstrated on an actual gasifier raw gas at low pressure, 0-10 psig. It is doubtful that this system can achieve high H_2S removal efficiencies at high pressure and its ability to handle other sulfur contaminants is yet to be demonstrated. Materials of construction for a commercial unit will also present a formidable problem.

Babcock and Wilcox Process--

This process is chemically similar to the Bureau of Mines process in that it utilizes iron oxide to remove H_2S from the gas at high temperatures. The difference lies in the material used by the two processes. While the Bureau of Mines' process uses a sintered material made from iron oxide and fly ash, the Babcock and Wilcox process starts out with carbon steel and generates an iron oxide scale on the steel surface which is then used as the desulfurization agent.

While the reactions are not pressure sensitive, the equipment lends itself to low (ambient) pressure operation. The desulfurizer uses a modified regenerative type air heater and is referred to as a "regenerative desulfurizer". Sulfur removal efficiency greater than 90 percent is projected. Regenerant gas is expected to contain 10 to 13 percent by volume SO_2. The process concept has been demonstrated on bench scale equipment and a hardware design has been developed.

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SECTION 6

DETAILED DESCRIPTION OF SELECTED GASIFICATION AND CLEANUP PROCESSES

INTRODUCTION

This section presents both a description of and the rationale leading to the selection of particular gasification and desulfurization processes for integration with a combined-cycle power system. The general approach was to select gasifiers that were representative of the various general types. These were combined with selected high- and low-temperature cleanup systems for comparison of the resulting integrated power plants.

SELECTION OF PROCESSES

Gasifiers

During this and earlier phases, at least one representative of each of the major gasifier types was selected for study. The coal gasifiers that have been used in defining the integrated power plants are:

1. Bureau of Mines Stirred Bed Gasifier
2. U-Gas Ash Agglomerating Gasifier
3. Koppers-Totzek Entrained Flow Gasifier
4. BCR-type, Two-Stage Entrained Flow Gasifier
5. Molten Salt Gasifier

Of these, all but the Koppers-Totzek process operate above atmospheric pressure. While pressurized operation presents a number of operational problems, it is a definite advantage when attempting integration with a combined-cycle power plant. This fact showed up in both the performance and cost of power from a plant using an oxygen-blown K-T gasifier. While the process is very desirable in all other respects, the need to compress the product gas resulted in a heat rate approximately 10 percent worse than the closest competitor. Also, the combined effect of performance and cost of air-separation and other additional equipment resulted in a cost of electricity nearly 30 percent greater than the competition. It was concluded that the process does not lend itself to use with combined-cycle power generation.

The Bu Mines Gasifier, which was selected as typical of the "Fixed Bed" type, does not permit recovery of the sensible heat in the gasifier effluent. This is due to the need to quench that stream and condense out the tars and oils that would otherwise foul the downstream equipment. In the current phase, emphasis was placed on overall performance improvement by the use of high performance power system technology. This includes a high temperature gas turbine with a 2400 psi reheat steam bottoming cycle. Efforts at integration of the Bu Mines gasifier with such a power plant showed them to be incompatible due to the lack of heat recovery from the raw fuel gas and the large quantity of steam required by the gasifier. It was concluded that the performance estimates contained in the previous report (Reference 6-1) are representative of the capabilities of systems using this particular gasifier. The estimated heat rates were 9200 Btu/kWh for low- and 8600 Btu/kWh for high-temperature cleanup. These can be compared with current estimates of 7900 and 7500 Btu/kWh with low- and high-temperature cleanup respectively for the advanced type gasifiers. Therefore, no further work was done on the power plants having a Bu Mines-type gasifier.

While the oxygen-blown, slagging Lurgi gasifier promises much better performance than the Bu Mines process, the only oxygen-blown gasifier that was used was the BCR-type two stage process. It was studied for comparison with its air-blown counterpart. Gasifier operating characteristics did not differ as widely and the effect of air separation and medium- vs low-Btu gas could be better appreciated.

For integration with a combined-cycle power plant, the important gasifier characteristics are:

1. No condensible hydrocarbons in effluent
2. Low gasifier exit temperature
3. Low steam feed rate.

The remaining gasifiers, U-Gas, BCR-type two-stage and the molten salt-type appear to be good selections on those bases. In terms of power plant heat rate, cost of electricity and emissions, the estimated performance of all three is the same within the accuracy of the study data.

Low-Temperature Desulfurization Processes

The majority of low-temperature desulfurization processes that were identified are commercially operative and could be used with both current and advanced gasification processes. In selecting those most applicable in treating coal derived fuel gas, the following factors were considered:

a. Sulfur removal capabilities, not only with respect to H_2S, but also other sulfur compounds such as COS and CS_2.

b. Selective absorption of sulfur compounds over carbon dioxide. Since CO_2 need not be removed from fuel gas intended for use in advanced power cycles, absorption of CO_2 represents an increased operating load on the system.

c. Type of absorbent insofar as the treated fuel gas may be contaminated by entrained or volatilized solvent which could be detrimental to downstream system components.

d. The system's tolerance to other contaminants present in the raw fuel gas such as ammonia, cyanides, phenols, and tars.

e. Overall energy requirements and operating cost.

A preliminary screening, reported in Reference 6-2, led to the selection of three processes for detailed comparison in an integrated system. These are the Benfield chemical solvent process and the Selexol and Rectisol physical solvent processes. This comparison involved a preliminary performance estimate of an overall integrated power system using each of the three processes.

For the purpose of this comparison, the entrained-flow BCR type gasifier was used under the following operating conditions:

Coal Type	
Illinois No. 6	
Feed, lb/hr	2000
Gasifier Operation	
Temperature, F	1800
Pressure, psia	500
Air, lb/lb, coal	3.422
Steam, lb/lb coal	0.567
Gasifier Production	
Net Gas, SCF/lb coal	74.53
Slag, lb/lb coal	0.087

Raw Gas Analysis, Volume %

N_2	CO	CO_2	H_2	CH_4	H_2S	COS	NH_3	H_2O
47.70	16.74	8.84	11.98	3.14	0.46	0.10	0.38	10.66

There is a net heat rejection from the gasifier of 0.384 MM Btu/hr which is available for high-pressure steam generation. Each desulfurization unit was designed to treat the raw gas to a residual sulfur content of 100 ppm. The acid gas from the regenerator was designed for a high (> 10 percent) H_2S concentration enabling the recovery of elemental sulfur via a conventional Claus plant.

A block flow diagram for the integrated cleanup systems is shown in Figure 6-1. Raw producer gas from the BCR gasification system at 1750 F is first cooled in the heat recovery section to about 300 F. The heat extracted is available for regenerative heating of the clean fuel gas, boiler feed water preheating, and steam generation. This section was optimized to give the highest overall plant efficiency for each case. The gas is further cooled below the dew point to about 120 F via direct water scrubbing which also removes most of the ammonia present in the raw gas. Sour water from this scrubbing operation is first steam stripped and the stripped gases are sent to an ammonia recovery section. The cooled, ammonia-free producer gas is desulfurized in the low-temperature acid gas removal section producing a clean fuel gas and a regenerated acid gas stream containing 13-22 percent hydrogen sulfide. Sulfur is recovered from the latter stream and from the ammonia recovery off-gas stream in a vapor phase Claus unit. A tailgas treating section is included and recovers a minimum of 99 percent sulfur.

Complete material balances for the Selexol Benfield and Rectisol processes in this application are given in Reference 6-1. Performance of the cleanup processes are summarized in Table 6-1 together with the utility requirements.

As expected for a chemical solvent system, the Benfield process requires 2.5 to 3 times more low-pressure steam for solvent regeneration than do the two physical solvent systems. This is partially offset by the higher power consumption required by the Selexol and Rectisol processes for mechanical refrigeration to obtain subambient operating tempratures.

Using these data, a preliminary analysis of the integrated power stations gave the following relative performance characteristics:

Cleanup System	Clean Fuel HHV Btu/SCF	Fuel Gas Regen. Temp. F	Gasifier Cold Gas Efficiency % Coal HHV	Overall Station Efficiency % Coal HHV
Selexol	142.6	685	73.4	31.2
Benfield	136.5	685	74.9	30.5
Rectisol	144.0	685	73.4	31.4

INTEGRATED LOW–TEMPERATURE CLEANUP SYSTEM

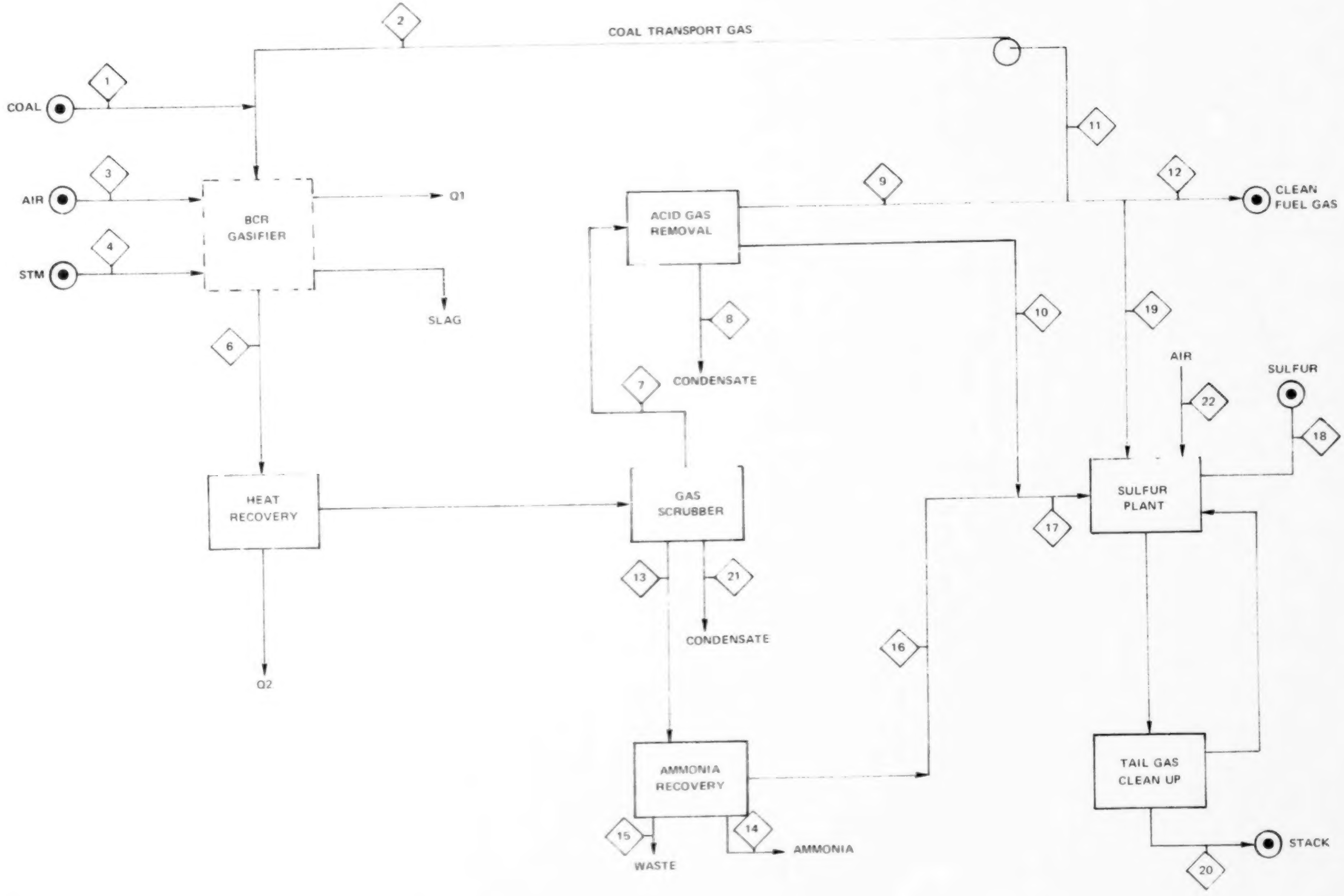

COAL TRANSPORT GAS
COAL
AIR
STM
1
2
3
4
BCR GASIFIER
Q1
SLAG
6
HEAT RECOVERY
Q2
ACID GAS REMOVAL
7
8
CONDENSATE
9
10
11
12
CLEAN FUEL GAS
GAS SCRUBBER
13
21
CONDENSATE
AMMONIA RECOVERY
15
WASTE
14
AMMONIA
16
17
19
AIR
22
SULFUR
18
SULFUR PLANT
TAIL GAS CLEAN UP
20
STACK
R1–19–4

FIG. 6–1

TABLE 6-1

SUMMARY OF LOW-TEMPERATURE INTEGRATED SYSTEMS

Process	Selexol	Benfield	Rectisol
Feed Streams(1)			
BCR Gas			
Flow, mph	426.45	426.45	426.45
T F	1750	1750	1750
P psia	450	450	450
Product Stream			
Sulfur			
Flow, lb/hr	74.9	74.9	76.2
T F	300	300	300
P psia	-	-	-
Transport Gas			
Flow, mols/hr	34.20	34.20	34.21
T F	300	300	300
P psia	550	550	550
Product Gas			
Flow, mols/hr	332.84	349.47	329.46
T F	100	250	90
P psia	430	430	430
Vol% N_2	54.70	52.32	55.20
CO	19.20	18.36	19.37
CO_2	8.66	7.02	7.88
H_2	13.73	13.14	13.86
CH_4	3.59	3.44	3.63
H_2O	0.11	5.66	
NH_3		0.06	0.06
H_2S	100 ppm	100 ppm	10 ppm
M. W.	25.47	24.81	25.31
HHV Btu/scf	142.6	136.5	144.0
Utilities			
Cooling Duty, MMBtu/hr	2.858	3.287	3.008
Steam @ 1300 psia, lb/hr	106.5	106.5	106.5
@ 65 psia, lb/hr	1020.0	2504.7	779.6
Electric Power, kw	60.8	25.5	41.8
Boiler Feed Water, lb/hr	219.2	222.8	223.1
Steam Condensate, lb/hr	1233.3	2718.3	994.8
Feed Gas Cooling, MMBtu/hr(2)	4.862	5.156	4.616

(1) Based on 2000 lb/hr Coal Feed to Gasifier

(2) Available for STM Generation and/or Gas Reheat

The efficiencies above do not necessarily represent the optimum power cycle configuration and fuel gas regenerative heating that can be achieved. However, in terms of their relative magnitudes for a particular system configuration, it can be concluded that the three low-temperature desulfurization processes can give comparable performance.

For a variety of reasons, including the relatively harmless solvent and moderate absorber temperature, the Selexol process was selected as the typical low-temperature process. Sulfur recovery is accomplished in a Claus plant with a Beavon unit for tail gascleanup.

High-Temperature Desulfurization Processes

Four high-temperature processes, the Conoco Half-Calcined Dolomite, Air Products Fully-Calcined Dolomite, Bureau of Mines Sintered Iron Oxide, and Battelle Molten Salt processes were compared on an integrated power plant basis in a manner similar to the low-temperature comparison. The comparison, complete with process flow sheets is reported in Reference 6-1. Results of the comparison are shown below:

Cleanup System	Clean Fuel HHV Btu/SCF	Clean Fuel Temp. F	Gasifier and Cleanup Cold Gas Efficiency % Coal HHV	Overall Station Efficiency % Coal HHV
Conoco	125.2	1610	76.1	36.0
Air Products				
Case 1	143.1	1550	53.8	29.1
Case 2	126.5	1630	73.7	35.5
Bureau of Mines	124.8	1000	72.5	31.6
Battelle	125.3	1610	76.1	34.9

For the Air Products fully calcined dolomite process, the large variation in performance is caused by absorption of CO_2 at the lower desulfurizer temperature. The reaction is exothermic. Thus, the bed must be cooled and re-calcination uses a like amount of energy.

The Bureau of Mines process suffers from the need to cool the raw gas down to 1000 F for desulfurization. Also, since the sulfur comes off the regeneration step as SO_2, the recovery of elemental sulfur is more difficult. However, it is ideally suited for the moving bed or Bu Mines-type gasifier for which it is being developed.

Of the two remaining systems, despite their relatively comparable performance, the Conoco process is preferred. The consequences of potential carryover of alkali metals are too severe to consider the use of a high-temperature molten carbonate system without a downstream water wash.

GASIFICATION PROCESS DESCRIPTION

For each process considered in this report, flow sheets and mass balances are given in Section 11. The coal composition given in Table 6-2 applies to all but the molten salt gasifier. While data for that process are based on an Illinois No.6 coal, the composition varies to some extent.

U-Gas Gasification Process

The U-Gas process is being developed by the Institute of Gas Technology. It is a fluidized bed system that can produce either low- or medium-Btu gas with either air- or oxygen-blown operation.

The use of a fluidized bed has many inherent advantages. In particular the bed acts as a catalyst for the gasification reactions and is expected to permit operation at relatively low temperature while completely gasifying the feed. The U-gas process is distinguished from other fluid bed processes in that it utilizes an "ash agglomeration" technique to concentrate the ash and remove it with minimum carbon content while operating the bed with a relatively high carbon content.

The gasifier is shown schematically in Figure 6-2 from Reference 6-3. The key to operation of the gasifier is the agglomeration and separation of the low carbon content ash from the bed. The U-gas gasifier accomplishes this and maintains a bed of approximately 70 percent carbon and 30 percent ash by proper design and operation of the grid and the fines return system in the bottom of the gasifier. The grid is sloped toward one or more inverted cones contained in the grid. Part of the fluidizing steam and air flow through the grid while the remaining fluidizing gas flows upward at high velocity through the throat at the cone apex. The ratio of steam to air in the fluidizing gas fed to the cone is chosen so that the resulting submerged jet in the cone is hotter than its surroundings. The temperature of the jet is maintained near the softening point of the ash particles for the specific coal being gasified. Ash particles preferentially stick together, and the agglomerates grow until they are heavy enough to move downward counter to the force of the gas stream in the apex of the cone. Thus, they fall out of the fluidized bed.

Fines elutriated from the fluid bed are separated from the product gas by two cyclones in series: the first inside the gasifier and the second outside. Fines removed by the first cyclone are returned to the bed by a standleg. Fines removed by the external cyclone are entrained in the inlet air/steam to the gasifier grid cone where they are instantly gasified because of the high temperature in that region.

With Illinois No.6 coal feed, the base case composition out of the gasifier is given in Table 6-3. All of the sulfur in the coal is assumed to appear as H_2S or COS and the two compounds are assumed to be at equilibrium with 94 percent of the total being H_2S. Ammonia is also

TABLE 6-2

FEED COAL COMPOSITION

Constituent	Weight %
Carbon	67.4
Hydrogen	5.1
Sulfur	3.8
Oxygen	9.6
Nitrogen	1.2
Ash	8.7
Water	4.2

Higher Heating Value - 12,200 Btu/lb

FIG. 6–2

ASH–AGGLOMERATING GASIFIER

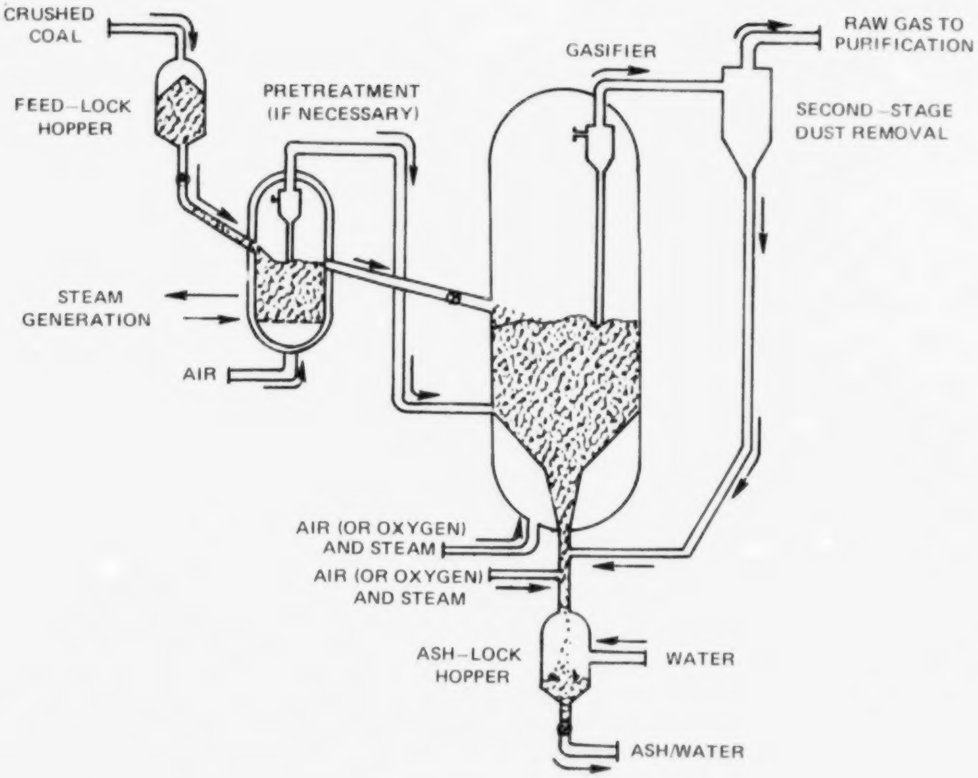

CRUSHED COAL
FEED–LOCK HOPPER
PRETREATMENT (IF NECESSARY)
GASIFIER
RAW GAS TO PURIFICATION
SECOND–STAGE DUST REMOVAL
STEAM GENERATION
AIR
AIR (OR OXYGEN) AND STEAM
AIR (OR OXYGEN) AND STEAM
ASH–LOCK HOPPER
WATER
ASH/WATER

TABLE 6-3

U-GAS GASIFIER EFFLUENT

	Mol Percent
CO	18.0
CO_2	8.47
H_2	15.42
H_2O	10.62
CH_4	3.03
N_2	43.82
NH_3	0.03
H_2S	0.59
COS	0.02

Mol. Wt 23.98

HHV - Btu/SCF 138.8
Cold gas efficiency (2) 80.1 percent
Steam/coal ratio 0.557
Air/coal ratio 3.01

assumed to appear in equilibrium concentration due to the catalytic effect of the fluid bed. The relatively low level would permit the use of high temperature cleanup without ammonia removal.

Raw coal is crushed to 1/4 in. size. The feed may contain up to 10 percent < 200 mesh material as generated in the crushing step. Noncaking, sub-bituminous coals or lignate can be fed directly to the gasifier from the crusher. Caking coals (Eastern bituminous, for example) must at present be pretreated by contact with air in a fluidized bed operating at gasifier pressure and 700 to 800 F. An oxidized outer layer forms on the coal particles, and this prevents agglomeration and possible blockage in the gasifier.

Heat evolved during pretreatment is removed by generating steam in heat-transfer coils which are immersed in the fluidized bed pretreater. Coal that has been pretreated is fed to the gasifier. Off-gases are fed to the bottom of the gasifier to destroy all tar and oils that evolve during the pretreating process.

The gasifier is a refractory lined, hot-metal-wall vessel. Steam is generated to provide cooling for the pressure vessel while the fluid bed reaction takes place at temperatures as high as 2000 F. System pressure can be as low as 100 psia (minimum level is determined by economics) but in this application is 400 psia, as determined by gas turbine pressure ratio and pressure drop in the fuel processing system and burner fuel distribution requirements. The operating conditions within the gasifier result in a product gas free of tars and oils. Thus no special cleanup procedures are needed.

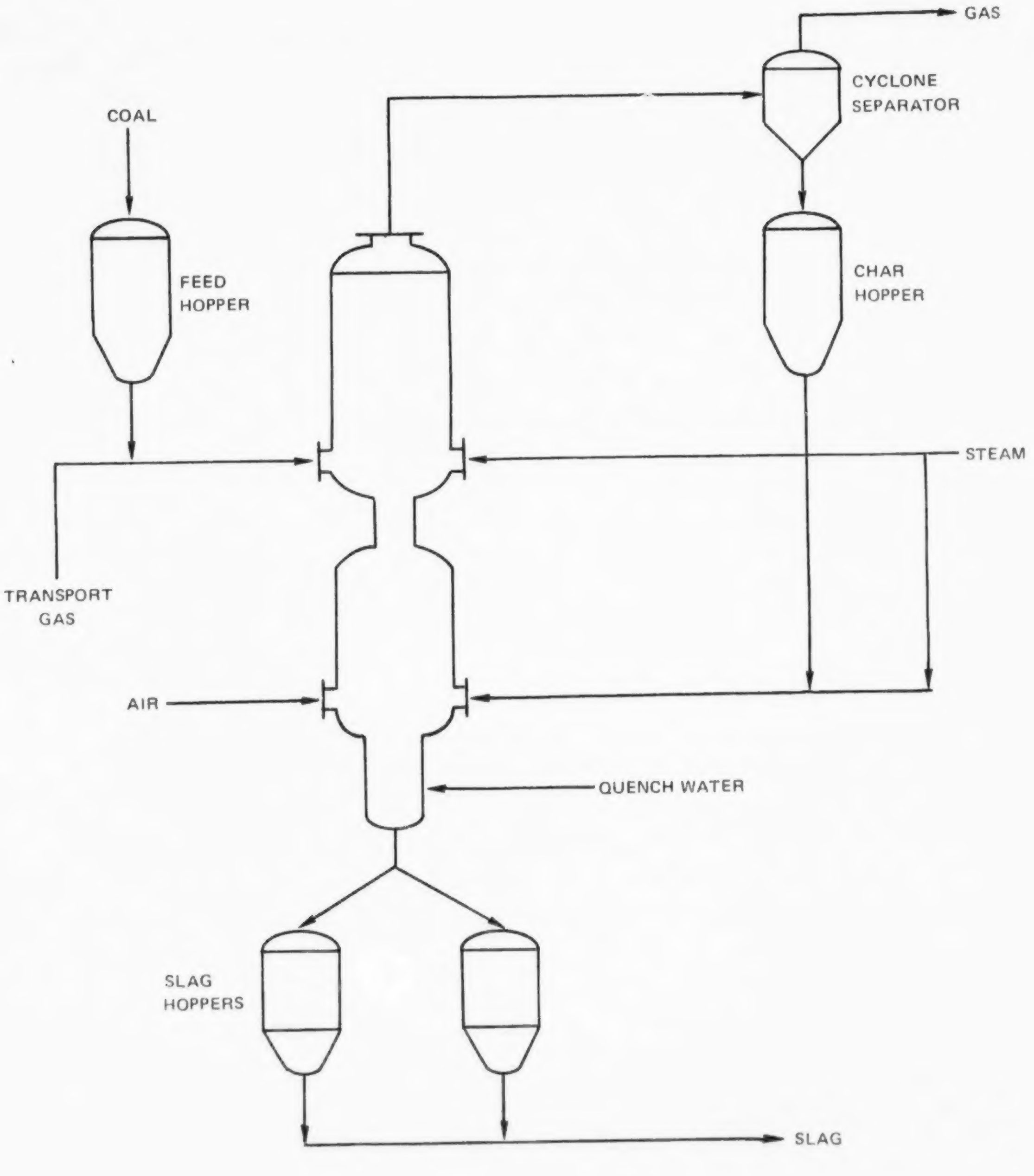
BCR Two-Stage Entrained Flow Process

The two-stage entrained flow gasifier developed by Bituminous Coal Research, Inc. (BCR) was selected as the second-generation gasification system for this study. An oxygen-blown, 120 ton/day pilot plant to produce 2.5 x 10^6 SCFD of SNG is under development at Homer City, Pennsylvania. Design of an air-blown version is currently underway at the Foster Wheeler Corporation.

A schematic diagram of the BCR gasifier with its auxiliaries is shown in Figure 6-3 (Reference 6-4). Run-of-the-mine coal is crushed, dried, and pulverized until 70 percent passes through a 200 mesh screen. The pulverized coal is metered from the feed hopper into hot transport gas recycled from the gas purification section and then fed into the upper stage of the gasifier. In this stage, the coal is entrained by the hot gases from the lower stage, is rapidly heated and devolatilized. At the high (1800 F) temperature, the products are expected to be methane, carbon monoxide, hydrogen and unreacted char. The gases leave the upper stage at approximately 1800 F.

FIG. 6-3

BCR ENTRAINED – FLOW GASIFIER


GAS
CYCLONE SEPARATOR
COAL
FEED HOPPER
CHAR HOPPER
STEAM
TRANSPORT GAS
AIR
QUENCH WATER
SLAG HOPPERS
SLAG

R1-19-3

Entrained residual char is removed from the gas by cyclone separators and recycled via superheated steam to the lower stage of the gasifier. Part of the char then reacts with steam and air at 2800 F to form synthesis gas and molten slag. The hot synthesis gas, containing unreacted char, flows to the upper stage for reaction with coal as described above. Molten slag collects and drains from the bottom of the lower stage into the slag pot where it is water quenched.

Overall, the gasifier reactions are endothermic and the process heat requirement is supplied by combustion of char with air. The air rate is regulated to maintain the operating temperature in the upper stage while the lower stage temperature is controlled by steam addition. Temperature in the lower stage is fairly critical since too high a temperature will damage the refractory and too low a temperature will cause the ash slag to freeze and accumulate.

For the current study, estimated performance of the gasifier was obtained from work done by Fluor under contract to EPRI. The estimate is based on minimum steam feed and is designed to improve performance of the integrated power plant. This is discussed in Section 8. Estimates for both air- and oxygen-blown operation have been given in Table 5-1. Note that while the air-blown case has a relatively low steam/coal ratio, that ratio is approximately 0.6 for oxygen-blown operation. This is due to the lack of the nitrogen diluent which helps to quench the reaction in the first stage. With pure oxygen, a significant amount of steam is needed to control the lower stage temperature.

The cold gas efficiency is 83 percent for air- and 85.7 percent for oxygen-blown operation. While data defining this gas composition are not available, it is expected that at the high temperature and with sufficient residence time, formation of tars, oils and phenols will be avoided. The resultant gas is expected to be quite clean relative to that encountered in the moving bed type of process.

Molten Salt Gasification

The molten salt gasification process is well suited to this application. It uses no steam and the catalytic effect of the melt, in addition to minimizing reactor volume, should tend to produce levels of ammonia and carbonyl sulfide that are close to equilibrium. Because a large part of the sulfur is retained in the melt, total sulfur compounds in the raw gas are at relatively low levels and at equilibrium COS would be on the order of 30 ppm. Other attributes of the process include the ability to handle caking coals and the ability to operate with either coarse or finely ground coal. Because both sulfur and ash are retained in the melt, a large part of the system is devoted to processing the melt to recover sulfur and regenerate the sodium carbonate. This process also incorporates a final fuel gas desulfurization step so that the use of a conventional desulfurization process is not required. Thus the process consists of several sections:

1. Gasification
2. Gas cleanup
3. Ash removal
4. Salt recovery

Gasification and Gas Cleanup--

In the gasifier, as indicated in Figure 6-4, coal is suspended in a bed of molten sodium carbonate and is gasified with air to produce low-Btu fuel gas. Coal mixed with recycled sodium carbonate is continuously charged to the gasifier, and the vigorous turbulence in the molten salt bed results in uniform distribution of coal throughout the bed. Air is injected into the molten bed and the partial oxidation of coal produces a low-Btu fuel gas free of coal ash. The ash is retained in the melt. Furthermore, it is also free of ammonia, tars, and heavy hydrocarbons. These constituents are destroyed in the molten sodium carbonate. Under conditions of low-Btu gasification, a high degree of sulfur retention as Na_2S results in the melt such that the raw fuel gas is low in H_2S. Trace amounts of alkali metals must be removed from the gas prior to firing so that a low-temperature scrub is necessary.

A number of complex chemical reactions, many of which involve catalytic interaction with the molten sodium carbonate, occur in the bed. Table 6-4 indicates, in simplified form, the most imporant of these. Partial oxidation, reaction (6-1), is the main route for gasification of carbon. If the air to coal ratio is too high, formation of CO_2 by reaction (6-8) becomes important. This is undesirable because the fuel gas quality suffers and the reactor tends to run hotter owing to the highly exothermic nature of reaction (6-8). In accordance with reactions (6-2) and (6-3) the volatile matter in the coal is broken down by the catalytic action of the molten salt and as a result the low-Btu fuel gas is free of tars and heavy hydrocarbons. It is reported that ammonia is also destroyed by the molten salt (Reference 6-5). Equilibrium calculations performed during this study did show some ammonia, however, it was at a relatively low concentration (180 ppm) and if not removed would not exceed allowable emission levels. The water-gas shift reaction, reaction (6-4), is maintained at equilibrium, and CO_2 and H_2O concentrations in the low-Btu gas product are low, being driven there by the reducing environment. Reaction (6-5) indicates an important feature of the process. Sulfur in the coal reacts with the melt to produce Na_2S. An equilibrium does exist between Na_2S in the melt and H_2S in the gas (reaction (6-6)). However, because CO_2 and H_2O concentrations are low, most of the sulfur is retained in the melt. Since H_2S levels are low resulting COS levels are very low (reaction (6-7)). Reactions (6-9), (6-10), and (6-11) are of minor significance in low-Btu gasification.

To remove ash and sulfur from the gasifier, a purge stream of melt is withdrawn and treated. Coal ash is separated, sulfur is removed as H_2S, and sodium carbonate is recovered for recycling to the gasifier. The ash level in the melt can be allowed to accumulate to as much as 30 percent, thereby minimizing the purge treating requirements.

LOW–BTU MOLTEN SALT COAL GASIFICATION PROCESS

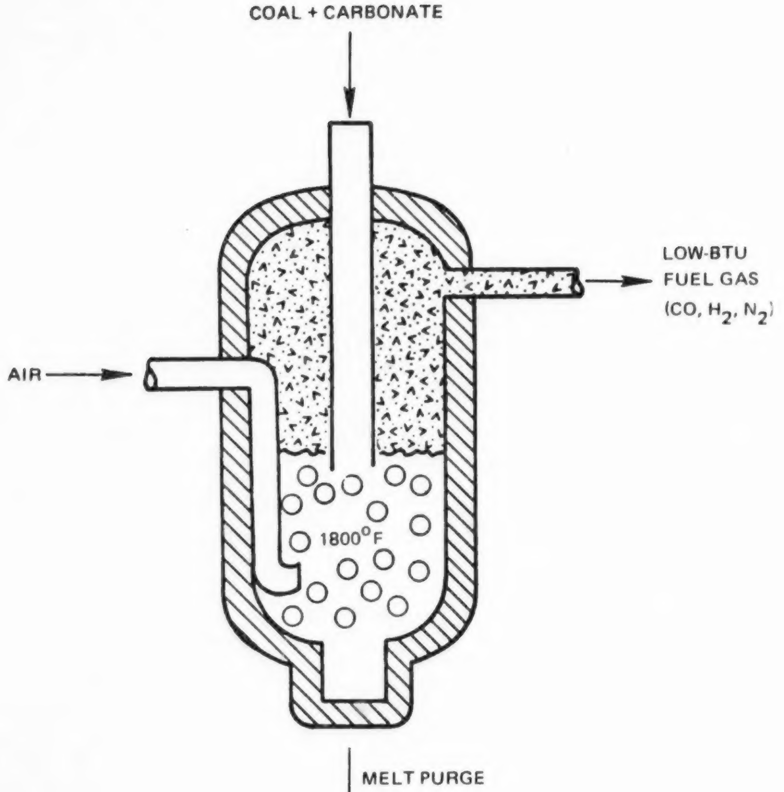

COAL + CARBONATE
LOW-BTU
FUEL GAS
(CO, H2, N2)
AIR
1800°F
MELT PURGE
(ASH, SODIUM CARBONATE, SULFUR)

TABLE 6-4

MOLTEN CARBONATE REACTIONS

$$C + 1/2\ O_2 \rightarrow CO \tag{6-1}$$

$$C_mH_n\ (\text{COAL}) \rightarrow \text{VOLATILES} + C \tag{6-2}$$

$$\text{VOLATILES} \quad CO + H_2 + CH_4 \tag{6-3}$$

$$CO + H_2O \rightleftarrows CO_2 + H_2 \tag{6-4}$$

$$S\ (\text{IN COAL}) + \text{MELT} \rightarrow Na_2S \tag{6-5}$$

$$Na_2S + CO_2 + H_2O \rightleftarrows Na_2CO_3 + H_2S \tag{6-6}$$

$$H_2S + CO \rightleftarrows COS + H_2 \tag{6-7}$$

$$C + O_2 \rightarrow CO_2 \tag{6-8}$$

$$C + H_2O \rightarrow H_2 + CO \tag{6-9}$$

$$C + 2H_2 \rightleftarrows CH_4 \tag{6-10}$$

$$Na_2CO_3 + H_2O \rightleftarrows 2\ NaOH + CO_2 \tag{6-11}$$

The operating conditions of the molten salt gasifier must be such that the bed is above the fusion temperature of sodium carbonate (1570 F), yet below the temperature where excessive vaporization of sodium carbonate and material corrosion problems arise. Normal operation is maintained between 1700 F and 1800 F. The temperature can be conveniently controlled by regulating the air to coal ratio. Since the process is basically one of partial oxidation, increasing the air to coal ratio increases the gasifier temperature at the expense of fuel gas heating value. The operating pressure is determined primarily by fuel gas pressure requirements. Pressure does not have a significant effect on the fuel gas composition. Table 6-5 shows the reference coal composition and the fuel gas composition resulting from gasification at 411 psia and 1800 F with 1000 F air.

To control alkali metals, two stages of gas cleaning are employed. In the first stage, a fluid bed cooler is used to quench the gas so the entrained and vaporized sodium salts condense and fuse on the surface of the particles in the fluid bed. By providing condensation sites while quenching the gas, the fluid bed prevents formation of a sodium carbonate or sodium oxide fume. In addition to removing vaporized and entrained sodium carbonate, the fluid bed cooler also serves to prevent carbon laydown. The reaction $2CO \rightarrow CO_2 + C$ is avoided by rapidly quenching the gas. At the same time a high level of recovery of the fuel gas sensible heat is achieved. In the second stage of cleaning, a water wash is employed to remove final traces of alkali metals.

Figure 6-5 shows a flow schematic of the gasification and gas clean-up sections in the plant. Ground coal mixed with makeup sodium carbonate is transferred to lock hoppers which alternately feed the gasifier. To remove ash and sulfur from the system, a purge stream of melt is withdrawn and sent to the ash removal and salt recovery section.

The raw fuel gas from the gasifier is quenched in the riser leg of the fluid bed cooler to 1200 F or lower. Contact with the large mass of solids in the fluid bed rapidly quenches the gas. Entrained or vaporized sodium salts are condensed and solidified on the surface of sand particles in the fluid bed. The fluid bed rises through a boiler where heat is removed and then flows into the disengager where the cooled fuel gas separates from the solids and flows out through a system of cyclones. To remove the trapped sodium carbonate from the fluid bed cooler, an intermittent purge is taken. The sand particles, now coated with sodium carbonate, are recycled to the gasifier where the sodium carbonate is recovered. The sand particles are removed in the gasifier purge. To replenish the fluid bed an intermittent charge of sand is required.

From the fluid bed cooler the clean fuel gas is further cooled by heat exchange with the clean gas in the recuperator. It then passes through an

TABLE 6-5

MOLTEN SALT GASIFIER
COAL AND RAW GAS COMPOSITION

Illinois No. 6

Carbon (wt. %)	59.62
Hydrogen (wt. %)	4.46
Oxygen (wt. %)	8.45
Nitrogen (wt. %)	1.00
Sulfur (wt. %)	3.10
Ash (wt. %)	10.37
Moisture (wt. %)	13.00
Higher Heating Value (Btu/lb)	10755

Raw Fuel Gas Composition

CO (vol. %)	28.33
H_2 (vol. %)	13.79
CH_4 (vol. %)	1.50
CO_2 (vol. %)	3.08
H_2O (vol. %)	2.35
N_2 (vol. %)	50.85
H_2S (vol. %)	0.10
Higher Heating Value (Btu/Scf)	152

GASIFICATION AND GAS CLEANUP

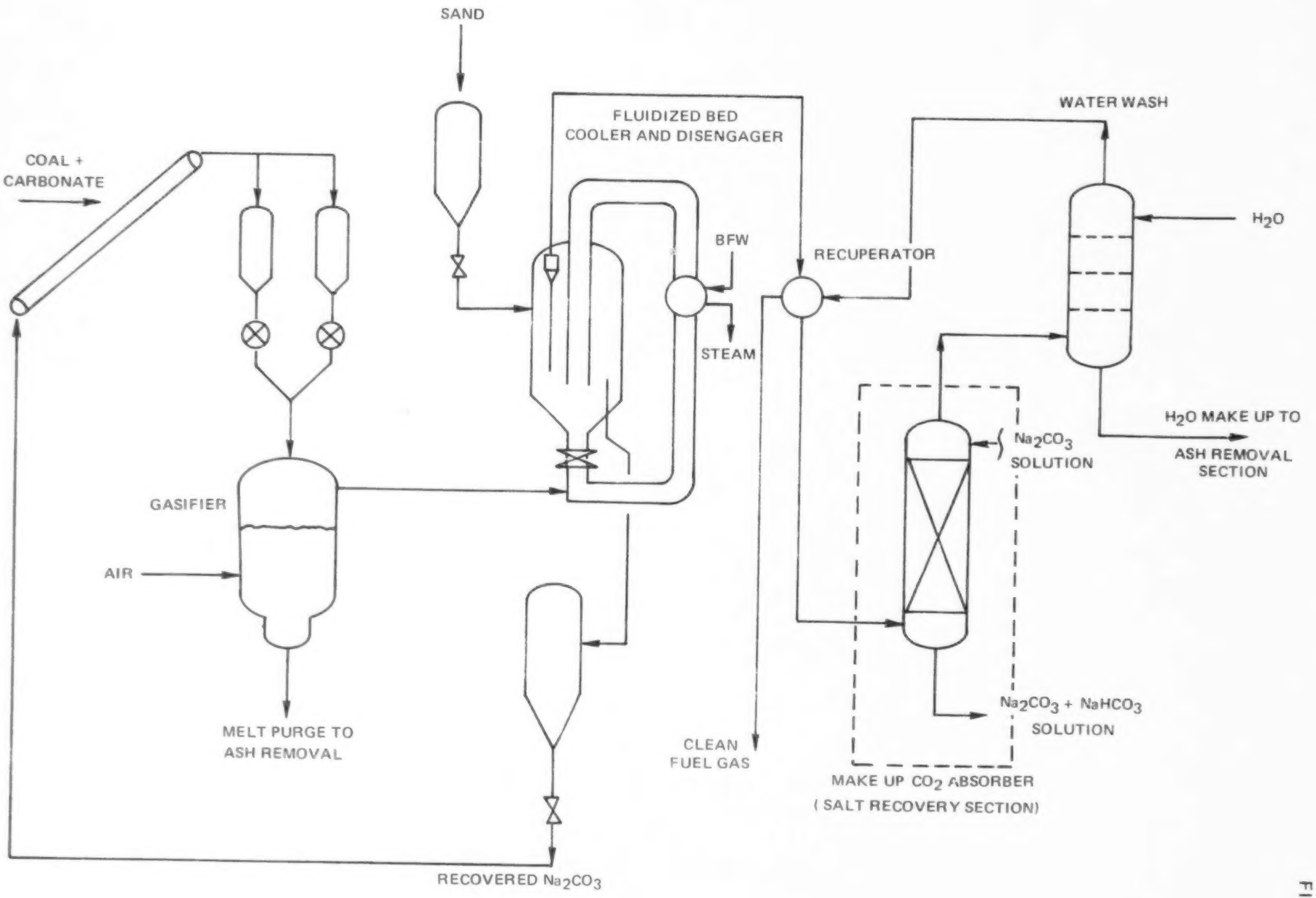

SAND
FLUIDIZED BED
COOLER AND DISENGAGER
WATER WASH
COAL + CARBONATE
BFW
RECUPERATOR
H_2O
STEAM
H_2O MAKE UP TO ASH REMOVAL SECTION
Na_2CO_3 SOLUTION
GASIFIER
AIR
Na_2CO_3 + $NaHCO_3$ SOLUTION
MELT PURGE TO ASH REMOVAL
CLEAN FUEL GAS
MAKE UP CO_2 ABSORBER
(SALT RECOVERY SECTION)
RECOVERED Na_2CO_3

FIG. 6--5

absorber where a portion of the CO_2 in the gas is absorbed to provide makeup CO_2 for the salt recovery section. The sodium carbonate solution in the absorber reacts with both CO_2 and H_2S thereby reducing the H_2S content of the gas stream from 1000 down to less than 200 ppm. Particulates carried out of the fluid bed cooler will be removed in the absorber. From the makeup CO_2 absorber, the fuel gas goes to the water wash which removes final traces of alkali metals. The clean fuel gas is then reheated in the recuperator and sent to the gas turbine combustor.

For the system studied here, the detailed stream compositions are presented in Section 11.

Ash-Removal and Salt Recovery--

The ash removal and salt recovery section treats the purge stream which removes ash and sulfur from the gasifier. By employing technology similar to that in practice in the pulp and paper industry, ash and sulfur are removed and the sodium carbonate is recovered. Figure 6-6 is a simplified flow diagram of the ash removal and salt recovery section.

The melt purge withdrawn from the gasifier at 1800 F and 400 psia is first quenched to 430 F by a circulating stream of sodium bicarbonate in the quench tank. The outlet stream from the quench tank flows to a hydroclone where solids are separated from the solution which is recycled around the quench tank. The quenched slurry is then flashed to produce 26 psia steam at 250 F.

From the flash tank, the slurry is sent to the dissolving tank where it is mixed with additional sodium bicarbonate solution and makeup water which includes all the process condensate collected during the process. The sodium salts are thus allowed to dissolve. The sodium sulfide and sodium hydroxide originally present in the melt purge are neutralized by sodium bicarbonate, according to the following reactions:

$$Na_2S + NaHCO_3 \rightarrow Na_2CO_3 + NaHs \qquad (6\text{-}12)$$
$$NaOH + NaHCO_3 \rightarrow Na_2CO_3 + H_2O \qquad (6\text{-}13)$$

The outlet stream from the dissolving tank is treated by a combination of clarification, filtration and washing of the solids to minimize sodium loss in the ash. Ash free filtrate solution is pumped to the H_2S stripper which uses the steam previously produced from the flash operation. The H_2S stripper is operated at low pressure (2.15 psia) and at low temperature (130 F) in order to minimize steam consumption. Along with the H_2S some CO_2 is also removed according to the following reactions:

$$NaHCO_3 + NaHS \rightarrow Na_2CO_3 + H_2S \qquad (6\text{-}14)$$
$$2NaHCO_3 \rightarrow Na_2CO_3 + CO_2 + H_2O \qquad (6\text{-}15)$$

ASH REMOVAL AND SALT RECOVERY SECTION

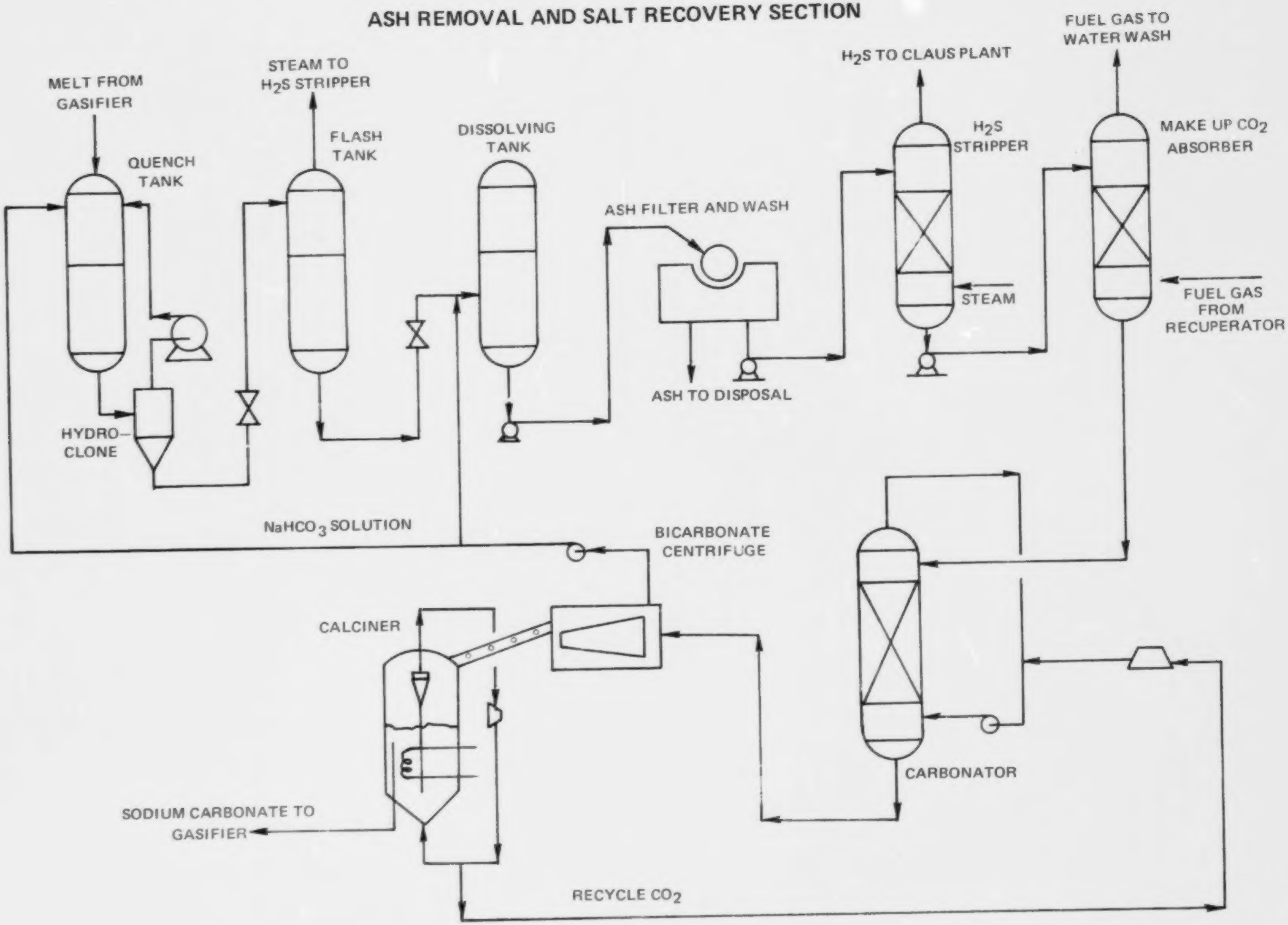

MELT FROM GASIFIER
QUENCH TANK
HYDRO-CLONE
STEAM TO H_2S STRIPPER
FLASH TANK
DISSOLVING TANK
ASH FILTER AND WASH
ASH TO DISPOSAL
H_2S TO CLAUS PLANT
H_2S STRIPPER
STEAM
FUEL GAS TO WATER WASH
MAKE UP CO_2 ABSORBER
FUEL GAS FROM RECUPERATOR
$NaHCO_3$ SOLUTION
BICARBONATE CENTRIFUGE
CALCINER
CARBONATOR
SODIUM CARBONATE TO GASIFIER
RECYCLE CO_2

FIG. 6–6

The overhead leaving the H_2S stripper at 130 F, containing water vapor, H_2S and CO_2, is cooled to condense and remove water vapor. The remaining gas stream (33% H_2S) is sent to conventional Claus plant for converting H_2S to elemental sulfur.

The stripped solution is reheated to 174 F and pumped to an absorber where makeup CO_2 and some H_2S are absorbed from the fuel gas according to the following reactions:

$$Na_2CO_3 + CO_2 + H_2O \qquad 2NaHCO_3 \tag{6-16}$$
$$Na_2CO_3 + H_2S \qquad NaHCO_3 + NaHS \tag{6-17}$$

This makeup CO_2 is required to compensate for CO_2 losses such as the CO_2 lost in the H_2S stripper and CO_2 lost in the gasifier by reaction of sodium carbonate with sulfur to form Na_2S. The makeup CO_2 absorber operates at the fuel gas pressure and at high temperature (175 F) in order to minimize the amount of water condensed from the fuel gas.

The solution is further carbonated with recycle CO_2 to precipitate solid sodium bicarbonate in the carbonation tower. The solution is cooled from 196 F to 140 F in the carbonation tower. At 140 F the sodium bicarbonate exceeds its solubility limit, and precipitates out of the solution.

The resulting slurry with 12% solids is separated by a combination of hydroclones and centrifuges. Sodium bicarbonate with 12% moisture is decomposed by indirect steam heating at 350 F in a fluidized bed calciner to produce sodium carbonate for recycling to the gasifier melt. The CO_2 produced during decomposition is recycled to the carbonation tower. The filtrate solution, along with condensate from other parts of the process is recycled to the slag quench and dissolving tanks.

DESULFURIZATION PROCESS DESCRIPTIONS

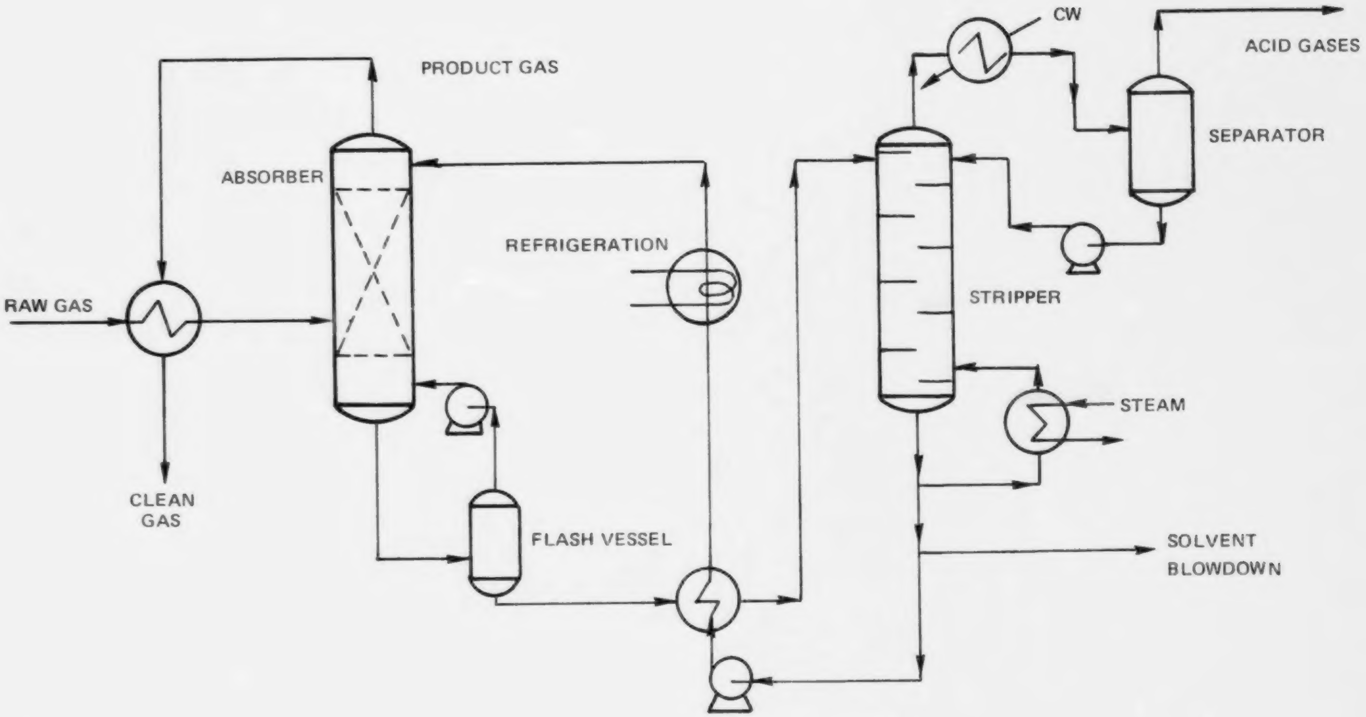
Selexol Process Description

The Selexol process, developed by Allied Chemical Corp., is a physical absorption process. It uses the Selexol solvent (dimethyl ether of polyethylene glycol) (Reference 6-6) to selectively remove H_2S in the presence of CO_2. Commercial applications currently include both selective absorption and bulk removal of acid gas.

For this application, a typical schematic is shown in Figure 6-7. Raw gas enters the absorber after having been cooled by heat exchange with the product gas. Where a high degree of COS removal is required, gas from the flash tank also enters at the bottom of the absorber. This is necessary to improve the process selectivity for COS with respect to CO_2. For the

FIG. 6–7

TYPICAL FLOW DIAGRAM–SELEXOL ACID GAS REMOVAL PROCESS


PRODUCT GAS
ABSORBER
RAW GAS
CLEAN GAS
REFRIGERATION
FLASH VESSEL
CW
ACID GASES
SEPARATOR
STRIPPER
STEAM
SOLVENT BLOWDOWN

78–01–287–3

basic systems currently under study, COS levels were not sufficiently high to warrant sizing of the system for any gases other than H_2S and the flash tank and recycle compressor were not required. To achieve very low emissions, systems designed for COS removal were also investigated.

Lean solvent, having been cooled by heat exchange with the rich solvent from the bottom of the absorber, is further cooled by refrigeration and introduced at the top of the absorber. Almost all of the H_2S is absorbed as the gas flows up through the tower. COS is less soluble and only 30 percent of that component is removed. Approximately 15 percent of the CO_2 is removed along with minor amounts of the other gases. The solvent vapor pressure is quite low (0.0002 mmHg at 60 F) (Reference 6-6) so that it does not contaminate the product gas.

The rich solvent is let down in pressure through a power recovery turbine and the absorbed gases are recovered by a combination of pressure letdown and stripping. Heat of absorption is low and most of the steam is needed to maintain the temperature differential between absorber and stripper. Acid gas to the Claus plant contains between 24 and 39 percent H_2S.

Other sulfur compounds, such as methyl mercaptan, carbon disulfide and thiophene are more soluble than H_2S and will be removed if present. The solvent is not degraded by impurities in the fuel gas. This, combined with the low vapor pressure, results in very low solvent makeup requirements.

Conoco Process Description

The process is shown in block diagram form in Figure 6-8. The raw gas passes through the fluidized bed desulfurizer where both H_2S and COS are removed. Dolomite is regenerated with a mixture of carbon dioxide and steam and the off gas sent to the sulfur recovery unit. Because the dolomite gradually becomes non-reactive, approximately 2 percent of the recirculating dolomite acceptor must be withdrawn and replaced. This is processed prior to discharge to convert the CaS to $CaCO_3$.

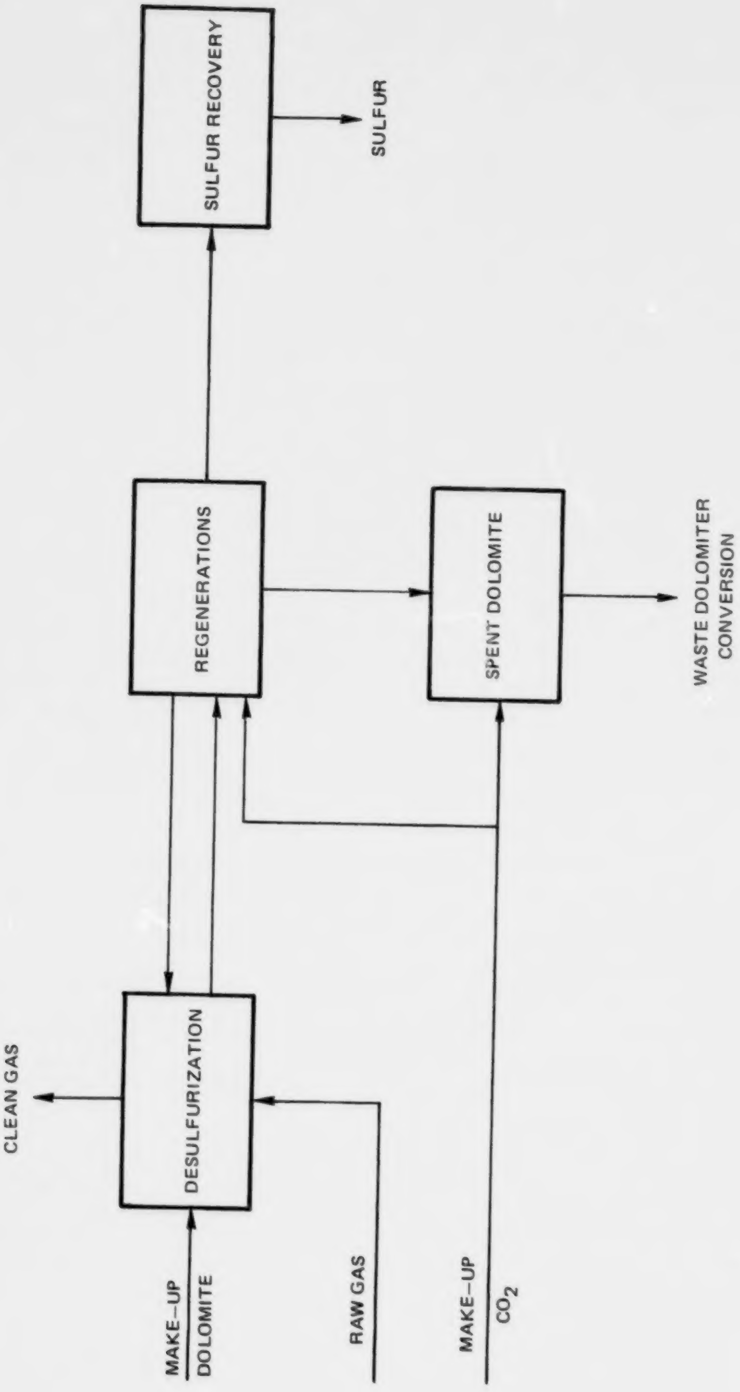
In addition to make-up dolomite, both steam and carbon dioxide are needed in the regenerator and spent dolomite converter. Steam is available from the bottoming cycle and the carbon dioxide must be separated from either the fuel gas or the gas turbine exhaust gas stream.

A schematic of the equipment arrangement and flow paths is shown in Figure 6-9. The process is described fully in Reference 6-7 and the systems used herein are based on the basic design presented in that report.

Gasifier product gas flows to the bottom of a fluidized gas desulfurizer. Most of the H_2S in the gasifier gas reacts with the $CaCO_3$ component of the dolomite as follows:

FIG. 6–8

CONOCO PROCESS BLOCK DIAGRAM



CLEAN GAS
DESULFURIZATION
MAKE–UP DOLOMITE
RAW GAS
MAKE–UP CO2
REGENERATIONS
SULFUR RECOVERY
SULFUR
SPENT DOLOMITE
WASTE DOLOMITER CONVERSION

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CONOCO HIGH – TEMPERATURE DESULFURIZATION

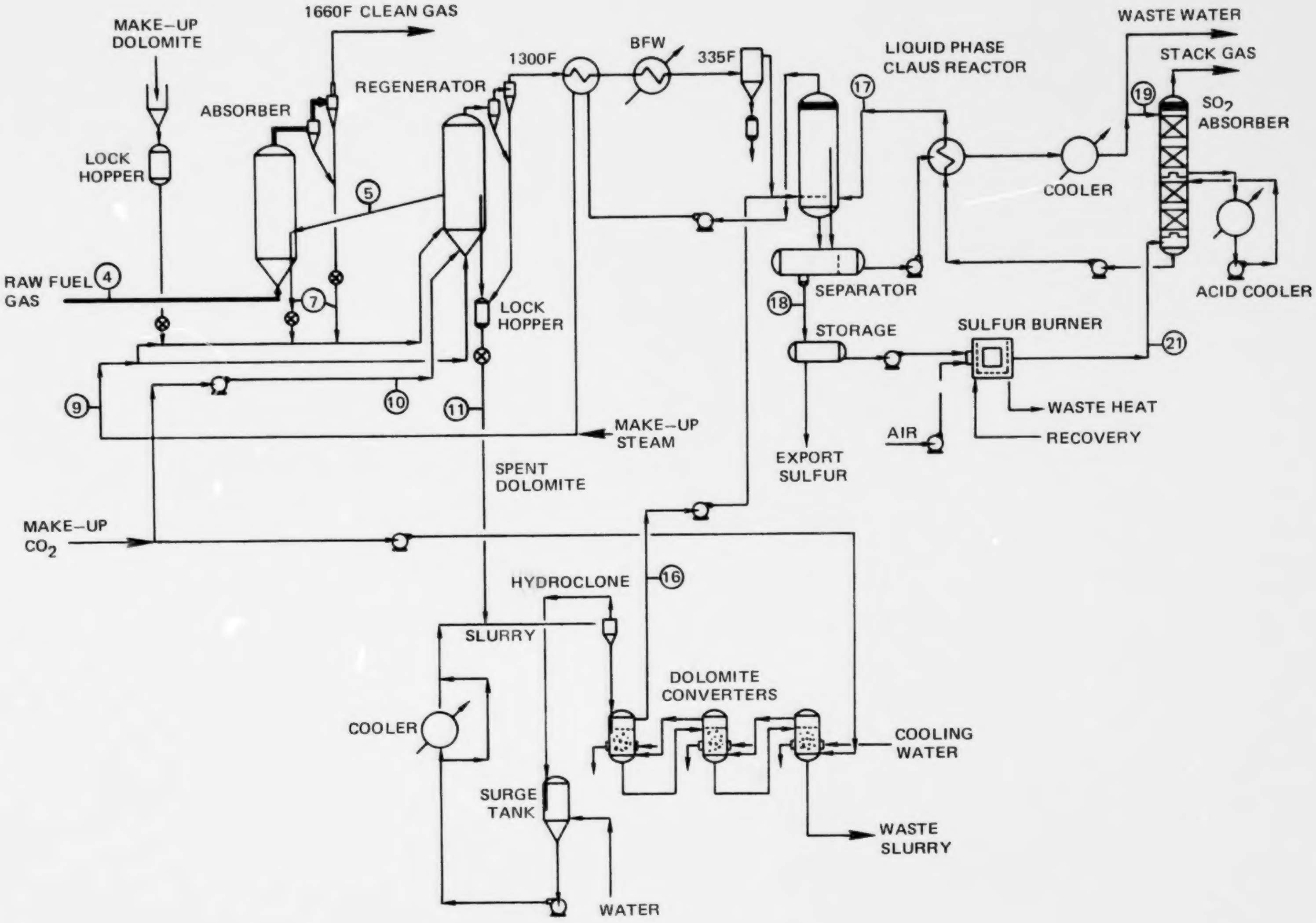

MAKE–UP DOLOMITE
1660F CLEAN GAS
LOCK HOPPER
ABSORBER
REGENERATOR
1300F
BFW
335F
LIQUID PHASE CLAUS REACTOR
WASTE WATER
STACK GAS
SO_2 ABSORBER
COOLER
ACID COOLER
RAW FUEL GAS
4
5
7
9
10
11
LOCK HOPPER
MAKE–UP STEAM
17
19
SEPARATOR
18
STORAGE
SULFUR BURNER
21
WASTE HEAT RECOVERY
AIR
EXPORT SULFUR
SPENT DOLOMITE
MAKE–UP CO_2
HYDROCLONE
16
SLURRY
DOLOMITE CONVERTERS
COOLER
COOLING WATER
SURGE TANK
WASTE SLURRY
WATER

FIG. 6–9

78–03–160–8

$$H_2S + MgO \cdot CaCO_3 \rightarrow MgO \cdot CaS + CO_2 + H_2O \qquad (6\text{-}18)$$

Although no data have been reported for the COS absorption high removal efficiencies are predicted according to the reaction:

$$COS + MgO \cdot CaCO_3 \rightarrow MgO \cdot CaS + 2CO_2 \qquad (6\text{-}19)$$

The desulfurized gas must then be cleaned of particulates and cooled prior to use in the combined cycle. This integration process is discussed in other sections of this report. The effect of gasifier operating conditions on the degree of desulfurization is discussed in Section 8.

The fluidized acceptor regenerator is maintained at 1300 F. The sulfided acceptor is recarbonated at these conditions by the reverse reaction,

$$MgO \cdot CaS + CO_2 + H_2O \rightarrow MgO \cdot CaCO_3 + H_2S \qquad (6\text{-}20)$$

The carbonated magnesium component of the make-up acceptor is also calcined at these conditions,

$$MgCO_3 \cdot CaCO_3 \rightarrow MgO \cdot CaCO_3 + CO_2 \qquad (6\text{-}21)$$

Spent acceptor (2 percent of the circulating flow) is withdrawn from the regenerators. This spent acceptor must be treated before disposal. Over 75 percent of the calcium component of this stream is in the form of CaS. If this were disposed of directly to the station ash pit, H_2S gas would slowly evolve as the CaS was hydrolyzed. To avoid this condition, the spent acceptor is directly contacted with CO_2 and water in three stages of stirred reactors to convert the CaS to $CaCO_3$. The overall reaction is,

$$MgO \cdot CaS + 2CO_2 + H_2O \rightarrow MgCO_3 \cdot CaCO_3 + H_2S \qquad (6\text{-}22)$$

Acid gas resulting from the acceptor stripping operation is passed through two stages of cyclones to remove entrained acceptor. The sensible heat content of this stream is then exchanged with recycle gas and with boiler feed water. Electrostatic precipitators are provided to remove entrained dust. The gas then flows to the bottom of the liquid-phase Claus reactor.

The concentration of H_2S in the gas from the acceptor regenerators is only about 3.6 volume percent. The liquid-phase Claus reaction developed for this process is uniquely suited to processing this gas. This was demonstrated during the experimental work described in Reference 6-8. Liquid sulfur is produced by the reaction,

$$2H_2S + H_2SO_3 \rightarrow 3\ S + 3\ H_2O \qquad (6\text{-}23)$$

Liquid sulfur and liquid water flow from the reactor to a decanter-type separator. Unreacted gas, saturated with water vapor at 310 F, is compressed and returned to the acceptor regenerator reactor. The sensible heat content of the liquid water from the separator is exchanged with the feed acid and charged to the SO_2 absorption tower. A slip stream of water is rejected to maintain water balance.

Approximately one-third of the sulfur produced is burned with stoichiometric air to produce SO_2 in the pressurized combustor. Excess heat is removed via cooling tubes in the walls. Water flowing down through the packed tower absorbs the SO_2 in the gas by,

$$SO_2 + H_2O \rightarrow H_2SO_3 \ \text{aqueous.} \qquad (6\text{-}24)$$

Most of the exothermic heat of reaction is removed by side stream coolers. The vent gas from the absorption tower is at 90 F and 205 psia. It probably would contain some residual SO_2 (assumed in this case as 0.3 volume percent).

Sulfur Recovery

In the case of the Conoco process, sulfur recovery is an integral part of the process. For low-temperature processes, where acid gas is produced with greater than 20 percent H_2S, the Claus process is used extensively. While it is a direct means of recovering elemental sulfur, recovery efficiency is generally in the 90 to 98 percent range and the tail gas can contain a significant amount of sulfur as H_2S, SO_2 or COS. Simple incineration of the tail gas is one possibility. In this study, the use of a Beavon unit for tail gas sulfur removal was elected. Both the Claus and Beavon units are described below.

Claus Sulfur Recovery

The basic Claus process was developed in the late nineteenth century. Since that time, many modifications have been made in the process by various engineering firms. The overall reaction that is the basis of the process is:

$$3H_2S + 1.5O_2 \rightarrow 3S + 3H_2O + 286{,}000\ \text{Btu} \qquad (6\text{-}25)$$

There are four principal variations of the process used in the United States (Reference 6-9):

Direct oxidation
Split flow
Straight through
Sulfur recycle

All are based on the same principle but find use depending on the acid gas concentration in the feed.

In the original Claus process, called "direct oxidation", H_2S was partially oxidized with air over a bauxite or iron ore catalyst in a single reactor. The reaction (6-25), which is highly exothermic, resulted in excessive temperatures and poor yields. This led to recycle of the tail gas for temperature control. It is now used for dilute acid feeds.

Two improvements were introduced in the 1930's by I.G. Farben. These are the "split flow" and "straight through" versions shown in Figure 6-10. In the "split flow" process, which is used here, one third of the H_2S is burned completely to sulfur dioxide in a waste heat boiler:

$$H_2O + 1.5O_2 \longrightarrow SO_2 + H_2O + 236{,}000 \text{ Btu.} \qquad (6\text{-}26)$$

The SO_2 is then reacted with the remainder of the H_2S over the catalyst:

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O + 50{,}000 \text{ Btu,} \qquad (6\text{-}27)$$

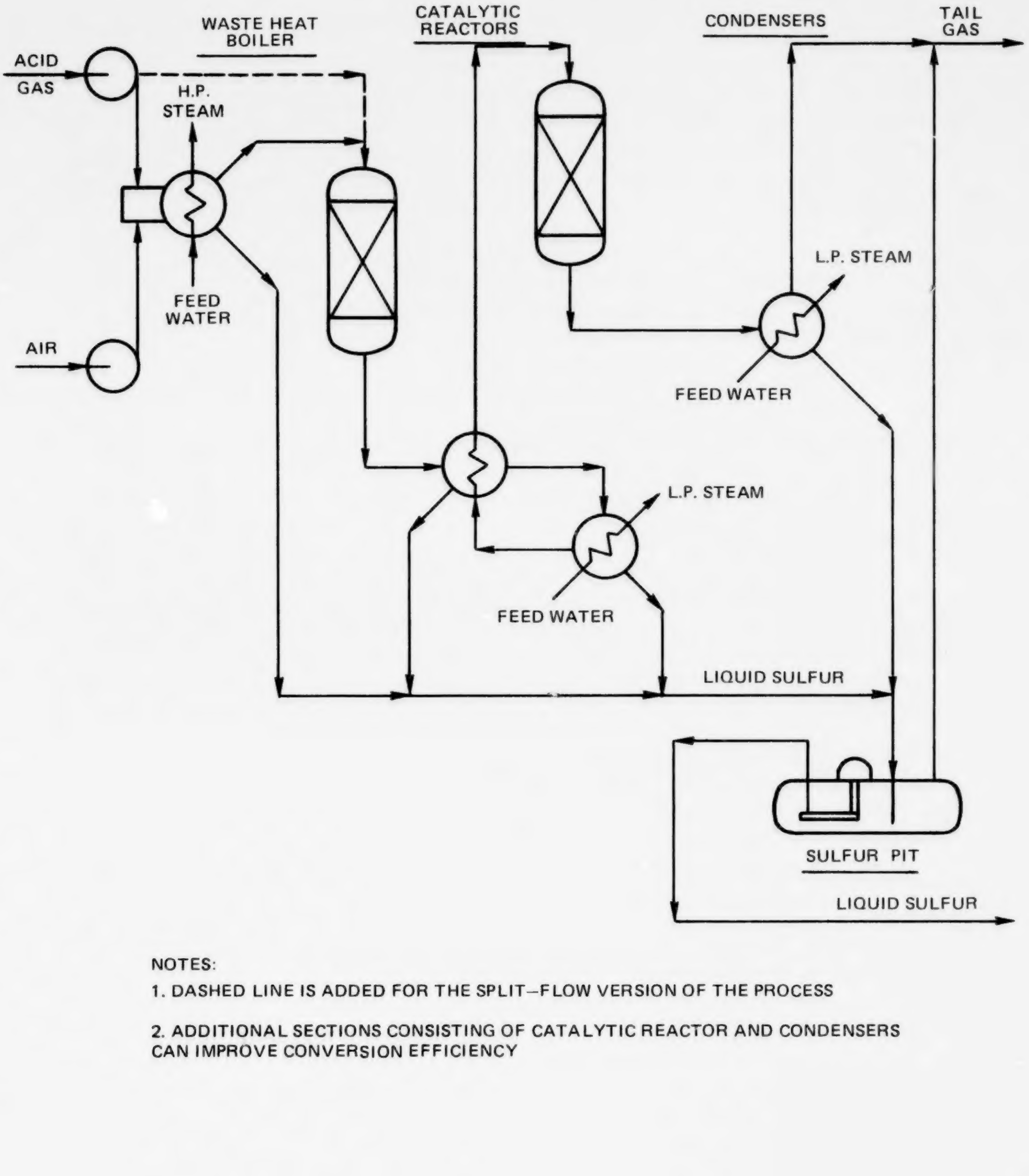
and less heat is involved.

In the "straight through" version, which is used at high (50 to 100 percent) H_2S concentrations, the acid gas is partially oxidized as in the direct oxidation process but in a waste heat boiler rather than over a catalyst. A large part of the conversion is achieved in the boiler and the cooled gas is sent to one or more catalytic stages for the desired conversion.

The acid gas streams feeding Claus plants sometimes contain so little H_2S that reaction temperatures cannot be maintained without supplemental heat. This gives rise to additional Claus plant variations, including addition of hydrocarbons to acid gas ahead of the burner and/or indirect preheating of the acid gas and/or air ahead of the burner or catalytic reactor.

One Claus plant variation, particularly useful for dilute acid gas having low sulfur oxide content, is called "sulfur recycle." In this, some product sulfur is burned with air to produce SO_2 and heat. The resultant stream is mixed with the acid gas in the appropriate ratio and sent to a

CLAUS SULFUR RECOVERY PROCESS


WASTE HEAT BOILER
CATALYTIC REACTORS
CONDENSERS
TAIL GAS
ACID GAS
H.P. STEAM
FEED WATER
AIR
L.P. STEAM
FEED WATER
L.P. STEAM
FEED WATER
LIQUID SULFUR
SULFUR PIT
LIQUID SULFUR

NOTES:

1. DASHED LINE IS ADDED FOR THE SPLIT–FLOW VERSION OF THE PROCESS

2. ADDITIONAL SECTIONS CONSISTING OF CATALYTIC REACTOR AND CONDENSERS CAN IMPROVE CONVERSION EFFICIENCY

catalytic reactor. This variation is applicable to H_2S concentrations of less than 15 percent.

Complete conversion to sulfur is prevented by equilibrium limits in the reactor. Improved conversion is achieved by adding reactor stages. The products of conversion are sulfur vapor and water vapor. They are condensed and removed between stages. Conversion efficiencies are generally in the mid-ninety percent range but up to 98 percent is attainable. The sulfur product is 99.9 percent pure and can be marketed.

Beavon Tail Gas Treating - This process consists of a hydrogenation step followed by a Stretford section for sulfur conversion. Named for D. K. Beavon, the process was developed by the Ralph M. Parsons Company. Final tail gas concentrations in the range of 40 to 80 ppm sulfur (as SO_2) have been reported for the process.

In the process, entering tail gas is mixed with hot flue gas as shown in Figure 6-11. Sufficient hydrogen is supplied, either with the tail gas or by fuel-rich combustion in the burner, to convert all the sulfur to H_2S. This is done over a cobalt-molybdate catalyst at temperatures between 550 and 750 F.

In the Stretford section, the H_2S is reacted with oxygen to form elemental sulfur and water. The process uses the following reaction sequence:

Absorber -

$$H_2S + Na_2CO_3 \longrightarrow NaHS + NaHCO_3 \qquad (6\text{-}28)$$

Reaction hold tank -

$$4NaVO_3 + 2NaHS + 2H_2O \longrightarrow Na_2V_4O_9 + 4NaOH + 2S \qquad (6\text{-}29)$$

$$Na_2V_4O_9 + 2NaOH + H_2O + ADA \longrightarrow 4NaVO_3 + 2ADA \text{ (reduced state)}$$

Oxidation tank -

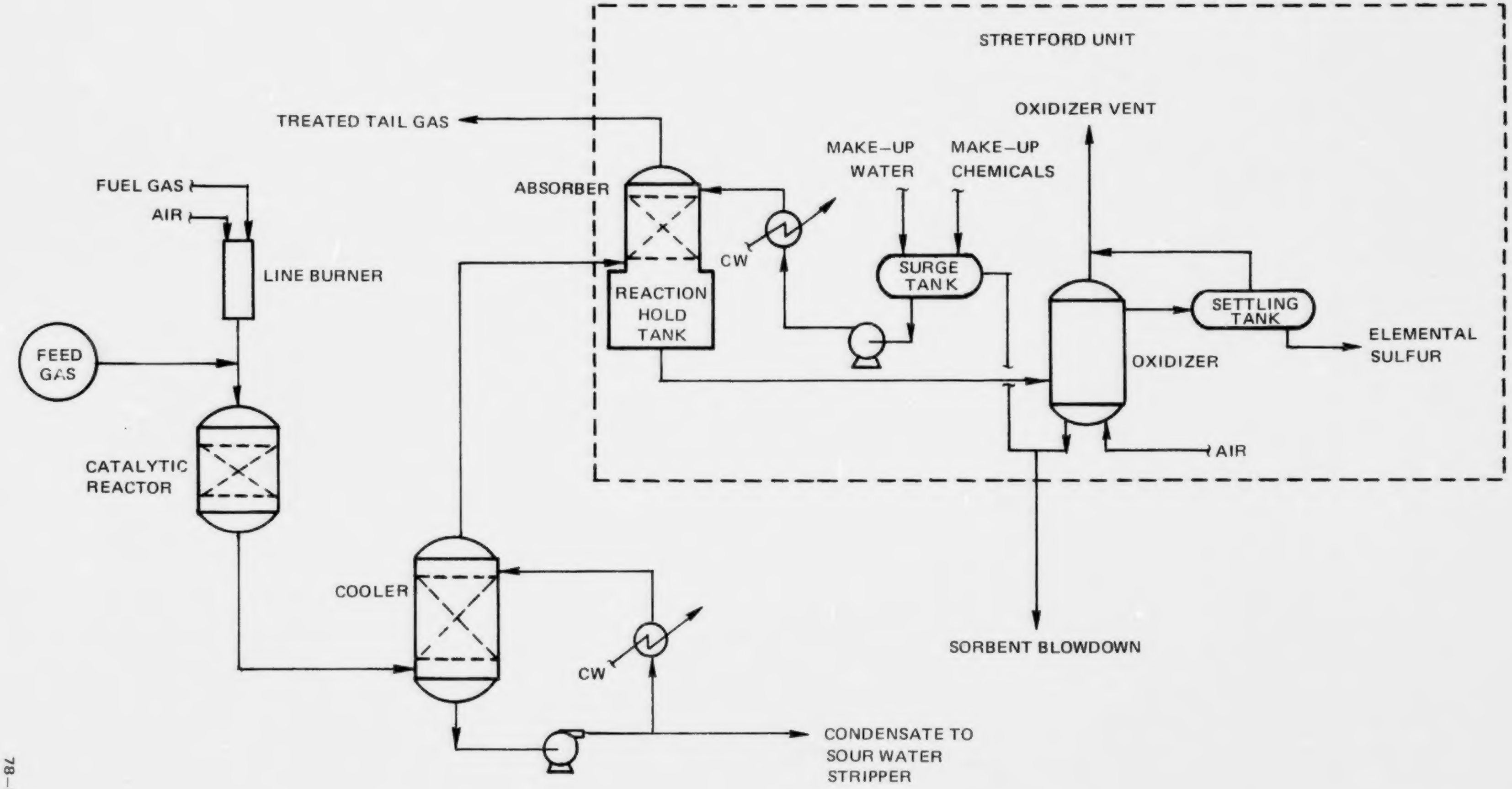
$$2ADA \text{ (reduced state)} + O_2 \longrightarrow 2ADA + H_2O \qquad (6\text{-}30)$$

In the absorber, H_2S reacts with sodium carbonate solution to form sodium bicarbonate and sodium hydrosulfide. The catalytic sodium vanadate solution reacts with the NaHS in the hold tank to form elemental sulfur. Air blowing with sodium anthraquinone disulfonate (ADA) regenerates the sodium vanadate catalyst. The ADA is regenerated by oxidation.

The overall effectiveness of the Beavon process is dependent on conversion efficiency of sulfur compounds to H_2S and subsequent efficiency of the Stretford unit. High conversion efficiencies are predicted for S, SO_2, COS and CS_2 (Reference 6-10). The Stretford is capable of reducing the H_2S content to less than 1 ppm.

FIG. 6–11

BEAVON TAILGAS CLEANUP PROCESS


STRETFORD UNIT
TREATED TAIL GAS
OXIDIZER VENT
MAKE–UP WATER
MAKE–UP CHEMICALS
FUEL GAS
AIR
ABSORBER
CW
SURGE TANK
LINE BURNER
REACTION HOLD TANK
SETTLING TANK
ELEMENTAL SULFUR
FEED GAS
OXIDIZER
AIR
CATALYTIC REACTOR
COOLER
SORBENT BLOWDOWN
CW
CONDENSATE TO SOUR WATER STRIPPER

78–01–287–2

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SECTION 7

COMBINED-CYCLE POWER GENERATION SYSTEMS

INTRODUCTION

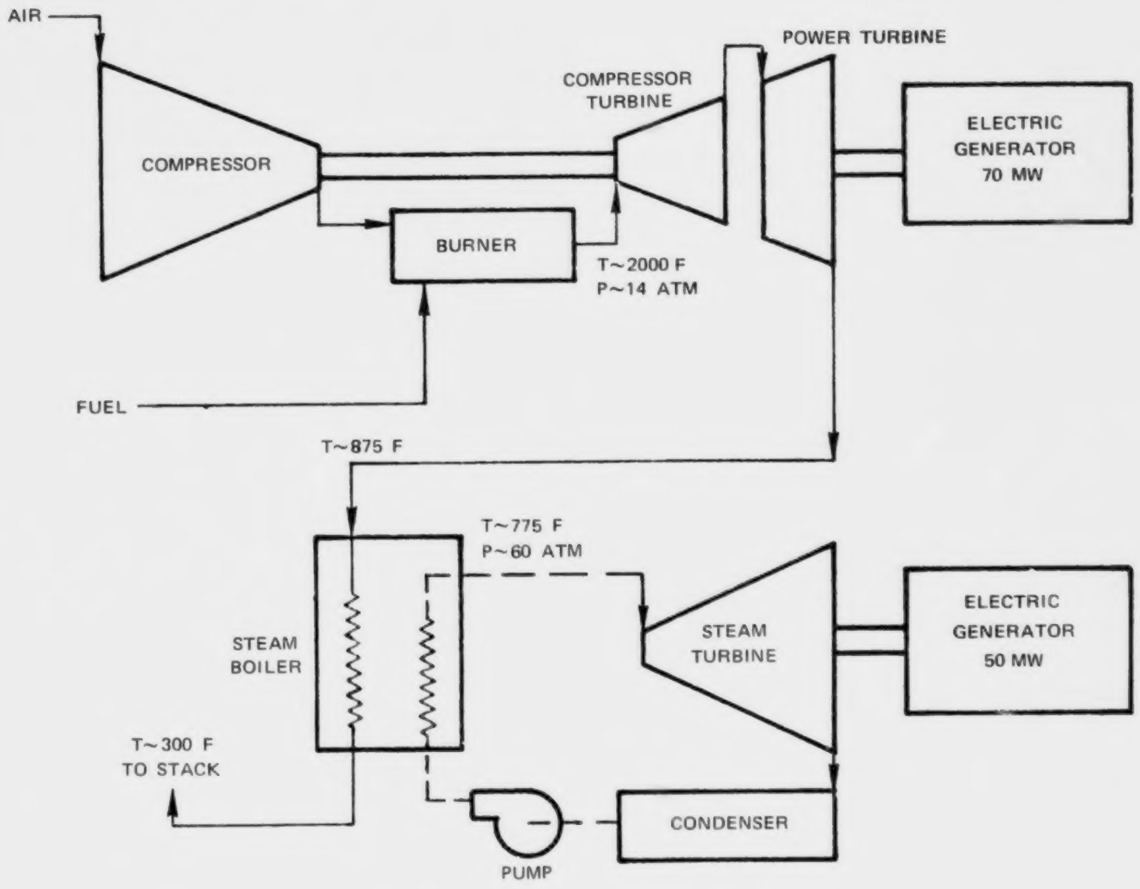
The term "combined cycle" can be used for a variety of systems where two cycles operating at different temperatures are interconnected. The heat rejected by the higher-temperature cycle is recovered and used in the lower-temperature cycle to produce additional power and improve overall conversion efficiency. The higher-temperature cycle is referred to as a topping cycle while the lower-temperature cycle is called a bottoming cycle. The combination used in this study is a gas turbine topping cycle with a steam bottoming cycle. It is currently the only combination of commercial significance.

A simplified schematic of the combined gas- and steam-cycle plant is shown in Figure 7-1. Exhaust gases leaving the gas turbine are typically at a temperature near 1000 F. This places an upper limit on the steam temperature generated in the waste heat boiler. As higher steam temperatures are generally associated with a higher steam cycle thermal efficiency, it is desirable to raise the gas turbine exhaust temperature. To do this would mean a decrease in the work extracted from the gas turbine cycle. Clearly, there is a tradeoff between gas turbine and steam cycle parameters. This trade-off is made even more complex by the addition of the gasification and cleanup processes. These present a source of heat to raise steam for the bottoming cycle. They also use steam that can be extracted from the steam cycle after doing some useful work.

The current study uses gas turbine and steam-cycle parameters that were developed as part of a DOE-sponsored study, the High Temperature Turbine Technology Program (HTTTP). As part of that program, parametric studies were conducted to determine the optimum combination of gas turbine inlet temperature, pressure ratio, metal cooling methods, steam cycle conditions and various other factors affecting overall power generating efficiency. All of these were evaluated in the context of a system comprised of a molten salt type gasification system integrated with a combined cycle power plant. This section presents a review of the process that led to the HTTTP selection of a combined cycle power plant having the following characteristics:

FIG. 7-1

WASTE-HEAT COMBINED GAS AND STEAM TURBINE SYSTEM


AIR
COMPRESSOR
BURNER
FUEL
COMPRESSOR TURBINE
T~2000 F
P~14 ATM
POWER TURBINE
ELECTRIC GENERATOR 70 MW
T~875 F
STEAM BOILER
T~775 F
P~60 ATM
STEAM TURBINE
ELECTRIC GENERATOR 50 MW
T~300 F TO STACK
PUMP
CONDENSER

Gas turbine pressure ratio	18:1
Gas turbine firing temperature	2600 F
Static parts cooling fluid	Water
Rotating parts cooling fluid	Pre-cooled air
Steam cycle pressure	2400 psi
Steam temperature	950 F
Reheat temperature	950 F

These same characteristics have been used with the exception of the cooling method for the static parts. The use of ceramic vanes which require no cooling has been assumed. This results in a very minor change in gas turbine performance. Previous phases of this EPA study had used a 24:1 gas turbine pressure ratio and non-reheat steam cycle. The reduced gas turbine pressure ratio results in exhaust temperatures which in concert with waste heat from the gasifier are able to support the high performance steam cycle. The result is a significant improvement in overall power generating efficiency.

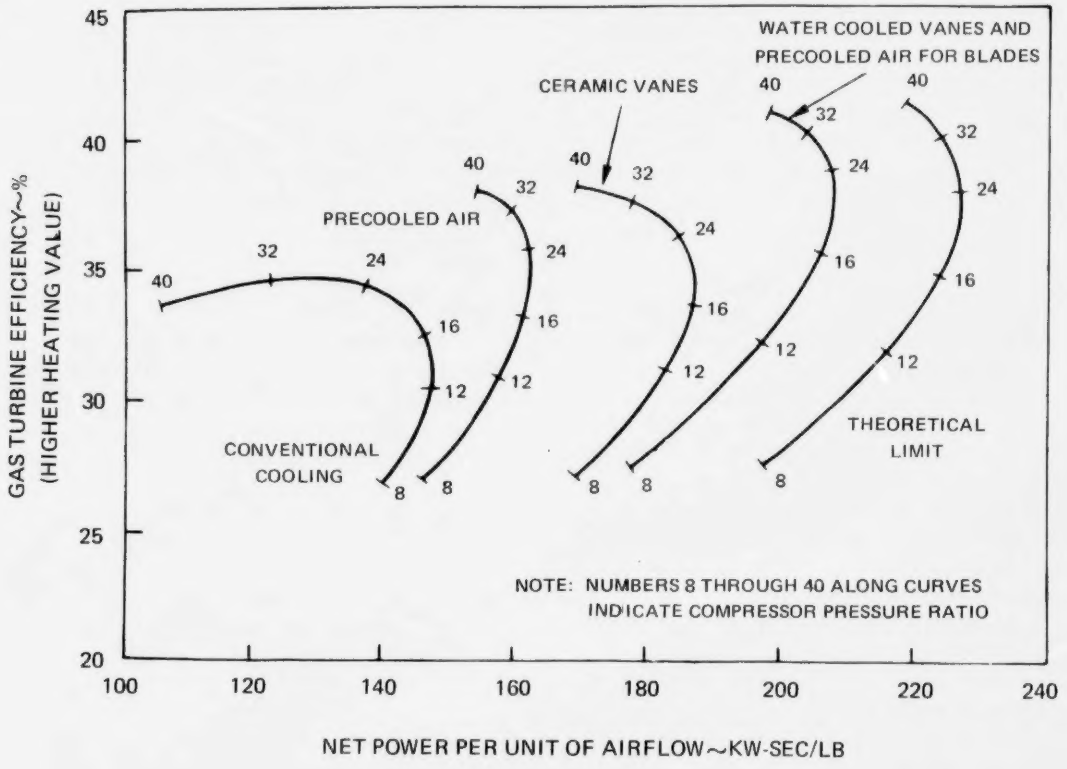
GENERAL PARAMETRIC ANALYSES

Analyses performed in earlier phases of this program and in the HTTT program were first directed towards defining the operating conditions of the gas-turbine engine. This was done by conducting trade-off studies that relate the interaction of engine pressure and temperature levels to performance levels and design requirements. Initially, these conditions were defined with respect to the engine alone, then modified to account for the interactions with the remainder of the system, thereby identifying a gas-turbine engine that provides an overall optimum system. For example, typical curves relating to the interactions among compressor pressure ratio and turbine cooling techniques with open-cycle gas turbine engine power output and efficiency are shown in Figure 7-2. This figure indicates that, while the relationship of performance and specific power is a strong function of cooling system type, there is a range of compressor pressure ratios generally between 12:1 and 24:1 which demonstrates a compromise between specific power, which is a measure of capital cost, and efficiency.

However, when the steam bottoming cycle is added to the design, the problem of optimizing the gas turbine becomes more complex. Two of the more important factors considered were the effects of the gas turbine operating point on steam system efficiency and the relative costs of the two systems. Increasing the compressor pressure ratio of the gas turbine while maintaining the defined turbine temperature has the effect of increasing the work output of the engine and decreasing the engine exhaust temperature. Although this augments the performance of the gas turbine cycle, it leaves less energy for the steam bottoming cycle. As a result, because the exhaust temperature is lower, the steam temperature is lower and the efficiency of the steam cycle is then lower. Thus, the goal is to select an operating condition that provides the highest overall performance.

FIG. 7–2

OPEN CYCLE GAS–TURBINE ENGINE PERFORMANCE TRADE–OFF CURVES



Definition of the steam bottoming cycle is dependent upon the waste heat available from the gasifier and cleanup system as well as that available from the gas turbine exhaust stream.

Analyses were performed to obtain the best match among these components for a definition of the best overall powerplant performance and cost. The two key items are the steam turbine and the waste heat boiler. The selection of operating conditions for these components seeks to ensure a sufficiently high rate of heat transfer by maintaining sufficient temperature difference between the gas turbine engine exhaust and the steam being heated in the waste heat boiler. The smaller this temperature difference, the larger and more costly the heat exchanger must be. In contrast, the larger the difference, the more overall efficiency decreases. These effects are discussed under thermal integration in Section 8. Selection of a "standard" combined-cycle power plant results in a slight penalty for some of the systems studied, but did not materially affect the overall conclusions.

On the basis of results acquired from general parametric studies, it is possible to identify the areas of interest concerning gas turbine and steam turbine operating characteristics. Figure 7-3 shows the generalized performance of simple-cycle and combined-cycle systems operating on liquid fuel, and demonstrates the expected effect with the integration of a coal gasification system. The range of interest defined for subsequent analyses is indicated by the shaded areas in Figure 7-3.

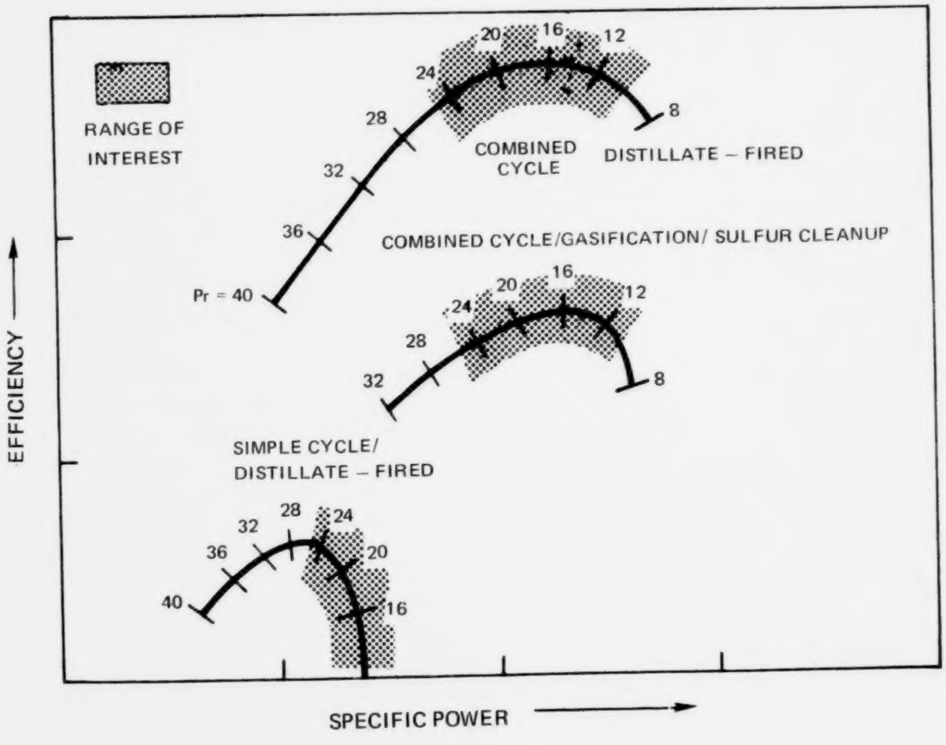
Operating Conditions

Gas turbine pressure ratio and firing temperature are the primary factors affecting overall system performance. However, the techniques used for cooling the high-temperature parts of the turbine and the allowable metal temperature have a direct bearing on performance at a given firing temperature. As part of the HTTT program for DOE, a metal temperature of 1600 F was selected as the maximum practical limit. Also, a cooling scheme that would make full utilization of existing technology was selected. It consists of water cooling for the static parts of the engine and the use of pre-cooled air for cooling the rotating turbine parts.

Using the molten salt gasifier, a parametric analysis was conducted. The results of this analysis indicated that an overall pressure ratio of 18:1 is nearly optimum for turbine stator inlet temperatures of 2600 F and 3000 F while a pressure ratio of 14:1 is nearly optimum for an operating temperature of 2400 F. Figure 7-4 presents the overall system efficiency versus net specific power as a function of the thermodynamic parameters of pressure ratio and turbine operating temperature. As indicated, as the turbine stator inlet temperature increases, the efficiency increases, but at a decreasing rate. The difference in maximum efficiencies between 2400 F and 2600 F is approximately one percent, whle the difference between 2600 F and 3000 F is 0.5 percent. In addition, the effect of pressure ratio on efficiency is less pronounced at the higher operating temperature levels. The difference between the optimum and lowest efficiency of the systems investigated is 1.5 percent at 3000 F, 2.8 percent at 2600 F, and 4.3 percent at 2400 F.

FIG. 7–3

TRENDS FOR COMBINED – CYCLE SYSTEMS



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HTTTP GAS TURBINE PARAMETRIC STUDY

Wa_{gt} = 815.6 LB/SEC

I_{SO} CONDITIONS

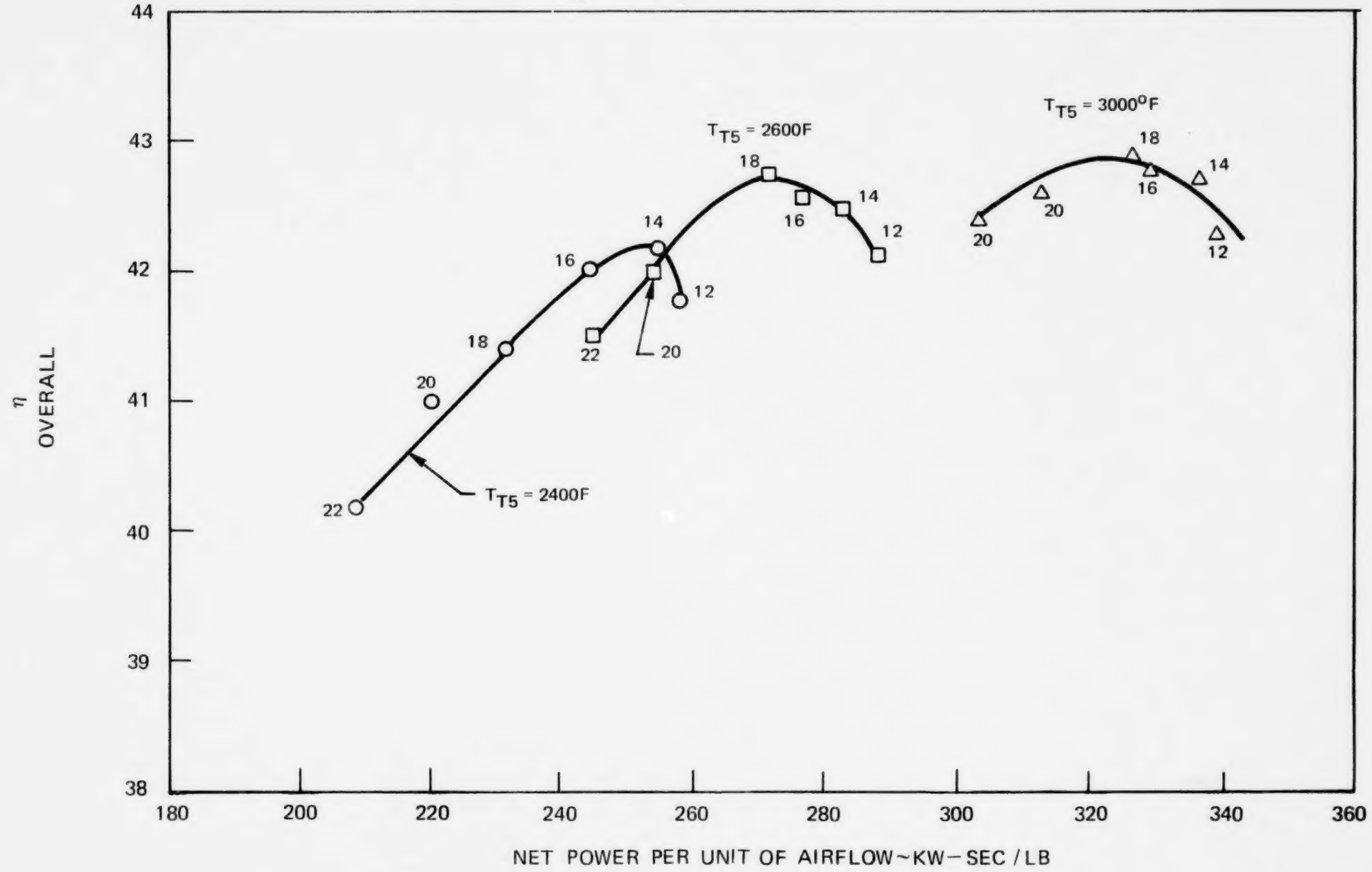

FIG. 7–4

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Using the data of Figure 7-4, a comparison of capital cost and heat rate can be made as shown in Figure 7-5. Taking cost and performance differences between a pressure ratio of 22 and 18 shows a capital cost increase of approximately $7.7 per kW as performance improves from a heat rate of 8224 to one of 7993 Btu/kWh. The capital expenditure to get such improvements in heat rate can be justified as coal costs exceed approximately $1.00 per million Btu. Because of the rapid trend in this direction, the higher efficiency appears to be desirable.

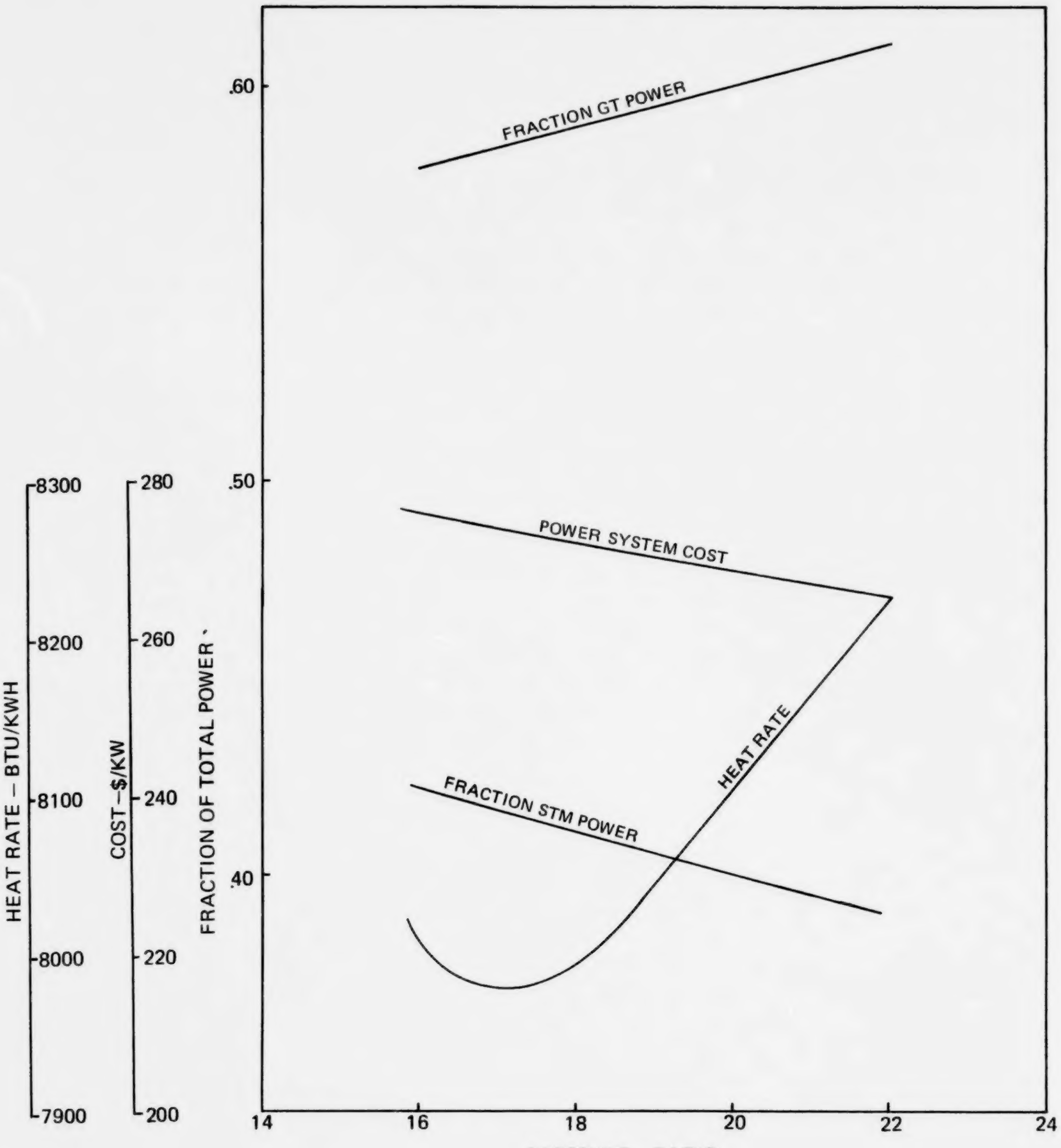
The costs for Figure 7-5 were developed using data from previous phases of the program. Power system costs were taken as $122 per gas turbine kW and $488 per steam turbine kW. Gasifier and cleanup costs are generally insensitive to pressure, at least over the range being considered here. For the pressurized equipment, theoretical relationships between pressure and vessel weight show weight to be independent of pressure. This would not hold true when comparing low- with high-pressure operation since at the low end, vessel or pipe thickness is dictated by considerations other than pressure. This is not the present case and the assumption of constant cost appears reasonable.

Steam Cycle Conditions

Having selected an 18:1 pressure ratio for the gas turbine, with the stator inlet temperature of 2600 F, the effect of changes in steam cycle conditions was examined. The results are shown in Figure 7-6. Gas turbine pressure ratios of 12 and 18 were considered. As expected, higher steam cycle temperature and pressure does improve overall plant performance. However, for the 18:1 pressure ratio, improvement over the 2400 psi/950 F/950 F reheat cycle was only marginal. Because that steam cycle is less demanding in terms of waste heat boiler design, it should make integration of the other gasification processes easier while not compromising performance. It was therefore selected for use in the integration studies.

FIG. 7–5

EFFECT OF CHANGES IN PRESSURE RATIO



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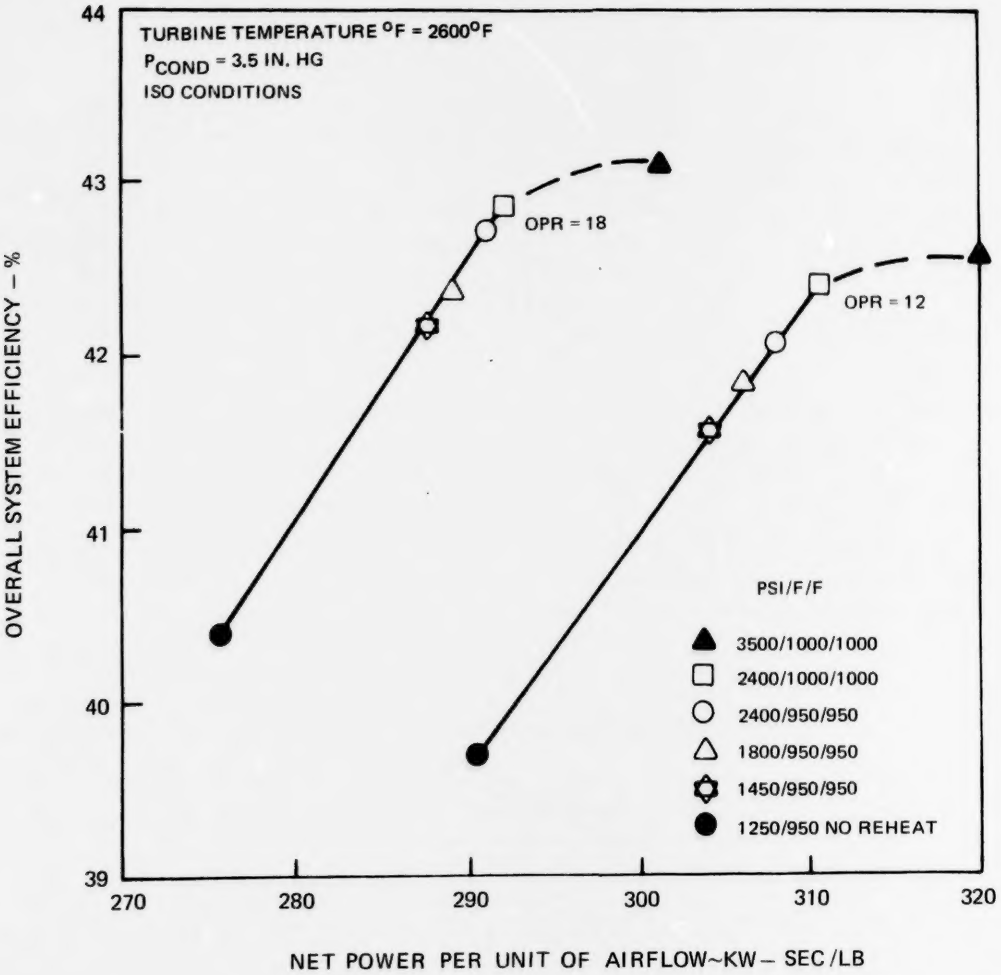
RESULTS OF STEAM CYCLE PARAMETRIC STUDY



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SECTION 8

INTEGRATION STUDIES

INTRODUCTION

Thermal integration can have a significant effect on overall system performance. Effective thermal integration simply means utilizing available thermal energy with minimum temperature difference between the source and the substance it heats. Thermal energy is available from the gasifier jacket, hot raw fuel gas, turbine cooling air, gasifier bleed air as well as from some of the process waste heat. Much of this heat is available at a temperature level suitable for raising relatively high-pressure steam. Moreover, process heat requirements generally involve the use of low-pressure steam which can be withdrawn from the steam cycle after having done some useful work.

When considering the complete system, the highest efficiency in the use of thermal energy is achieved when it is converted at combined-cycle efficiency. This means using high-temperature cleanup or reheating the fuel gas after a low-temperature cleanup. A number of thermal integration studies, including the effect of fuel gas reheat, were conducted during the earlier phases and are summarized in this section. In general, the results of those studies have been applied in defining the current integrated system schematics used for determing performance and environmental intrusion. In the case of clean fuel gas temperature, a maximum limit of 1000 F was established to allow the use of a pre-mix combustor for NO_x control. While this effectively limits some of the advantages of high-temperature cleanup, it appears to be essential if NO_x emission levels are to be within allowable limits.

Besides thermal integration, other integration studies were conducted to determine the effects, especially environmental effects, of operating temperature on the Selexol system and of steam feed rate on the U-Gas gasifier. Also examined was the effect of achieving equilibrium in each of the gasifier effluents. This last study was intended to identify potential species and concentration levels that could be expected, at least on the basis of equilibrium predictions.

The Selexol solvent is quite sensitive to operating temperature level. In natural gas sweetening plants, absorbers generally run at subambient temperatures. Because of the utility industry's desire to avoid the complexity of a refrigeration system as part of the process plant, a study of operating temperature was made. For the degree of sulfur removal considered in this study, refrigerated operation showed a decided advantage. Two different levels of sulfur removal were considered and the advantage of refrigerated operation was seen to increase with increased sulfur removal.

Earlier phases of this study had shown the effect of reduced steam feed to the gasifier. This improved cold gas efficiency and reduced the heat lost by condensation of water vapor in the raw fuel gas when it is cooled prior to desulfurization. Data used in this study for the BCR-type gasifier showed the lower steam feed rate (0.15 lb steam per lb coal vs. 0.57 lb steam per lb coal) resulted in a change in cold gas efficiency from 78.5 percent to 83 percent. This reduced steam feed rate also results in a lower air/coal ratio and a lower concentration of both CO_2 and H_2O in the raw gas. These low concentrations tend to improve performance of the high-temperature Conoco desulfurizer.

For the U-Gas process, the effect of steam feed rate on desulfurizer performance was determined and the point of minimum practical steam feed (based on discussions with IGT) was used in the subsequent system evaluation. Performance improvements (approximately 3%) for the U-Gas system with low steam feed were also identified for low-temperature sulfur removal.

Gasifier modeling efforts of previous phases were continued. Considering the goals of this study, the most important part of the modeling effort was the adaption of the chemical equilibrium calculation procedure to model the gasification process. Use of this tool allowed equilibrium limits to be defined for the formation or reduction of potential pollutants.

THERMAL INTEGRATION

The gas turbine operating parameters and steam cycle conditions have a significant effect on the overall thermal integration. Previous studies, for example, considered the use of a supplementarily fired waste heat boiler as a means of improving overall system performance. The idea was simply to raise the gas temperature in the boiler. This, in turn, allows improved steam conditions which increase steam cycle efficiency and improves the efficiency of utilization of waste heat. Overall efficiency is given by:

$$\eta_{cc} = \frac{\eta_{gt} + \left[(1 - \eta_{gt}) + \frac{W_s}{W_f}\right]\left(\frac{T_{ex} - T_{st}}{T_{ex} - T_a}\right) \eta_s}{1 + \frac{W_s}{W_f}} \qquad (8\text{-}1)$$

where

η_{cc} = combined cycle efficiency

η_{gt} = gas turbine efficiency

η_s = steam cycle efficiency

W_f = primary fuel flow

W_s = supplementary fuel flow

T_{ex} = gas turbine exhaust temperature

T_a = ambient temperature

T_{st} = stack temperature

A high turbine firing temperature combined with utilization of gasifier and raw fuel gas thermal energy make it possible to utilize a high-performance steam cycle without resorting to supplemental firing.

When there is no supplemental firing of the waste heat boiler, the equation simplifies to:

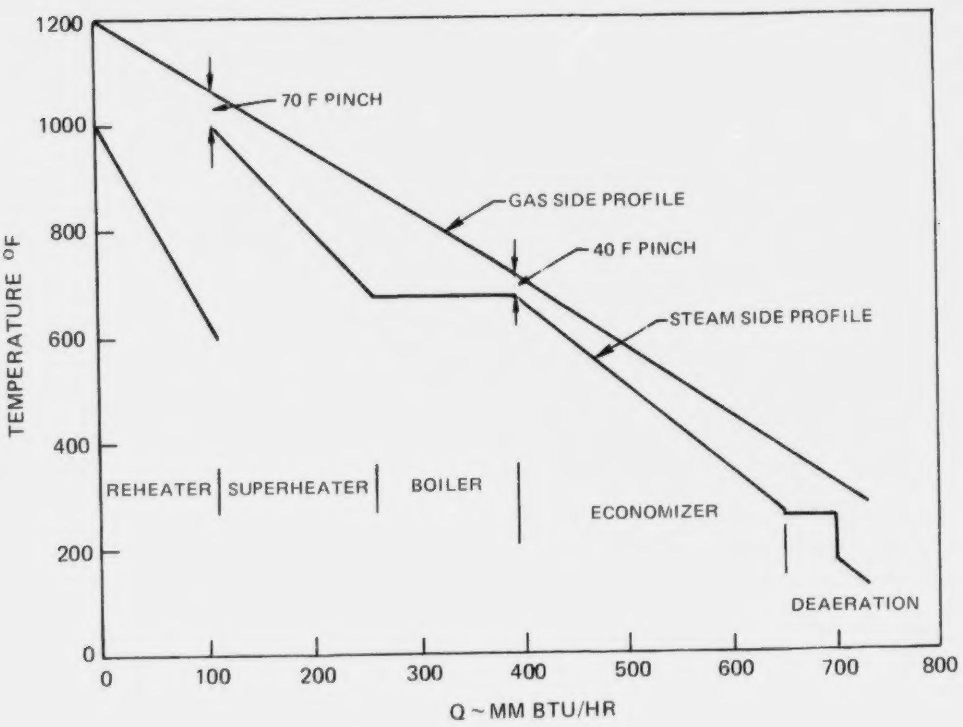
$$\eta_{cc} = \eta_{gt} + (1-\eta_{gt})\ \eta_s \frac{T_{ex} - T_{st}}{T_{ex} - T_a} \qquad (8-2)$$

and the importance of steam cycle efficiency is quite apparent, all other things remaining equal.

In each of the systems studied, virtually all steam is raised in the gasifier jacket and the hot fuel gas boiler. The gas turbine waste heat boiler is used almost exclusively for superheat, reheat and economizing. A typical Temperature Vs. Heat Flow (T-Q) curve for the waste heat boiler is shown in Figure 8-1. It shows the problems involved in utilizing the thermal energy of the turbine exhaust while maintaining minimum temperature differentials consistent with high-cycle efficiency. Because of the varying amounts of heat available from the gasifier, fuel gas stream, and other sources, waste heat boiler designs vary considerably among the various systems.

In the case of the oxygen-blown BCR type system with Selexol cleanup, the waste heat boiler is quite complex. In that system, it would be necessary to raise some steam at reheat pressure in the waste heat boiler to achieve a 300 F stack temperature which was set as a lower limit for all systems. Modification of the steam conditions could eliminate some of the

TYPICAL T–Q DIAGRAM FOR WASTE HEAT BOILER



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complexity; however, further iteration of the design was judged unwarranted as it would not materially change the study results. All heat recovery systems were able to be designed with a minimum superheater pinch of 70 F and boiler pinch of 40 F.

An important factor that does not appear in Equation 8-2 for combined-cycle efficiency is the effect of fuel gas temperature. The sensible heat associated with the fuel gas can be converted at full combined-cycle efficiency if the gas were reheated after cleanup and prior to being burned. The alternative to reheating would be to use that same heat in the steam cycle. Efficiency of the steam cycle is significantly lower ($\eta_{stm}/\eta_{cc} \sim 0.6$) than the overall combined-cycle efficiency.

The effect of reheating the clean fuel gas by heat exchange with the hot raw gas in a BCR system with Selexol cleanup is shown in Figure 8-2. As clean fuel temperature to the burner is increased, overall plant efficiency is improved at the cost of an increase in heat transfer surface area. The performance improvement depends heavily on the temperature level below which sensible heat in the hot fuel gas stream is not recovered (i.e., minimum temperature at exit of boiler or hot side of fuel regenerator). Recovery of this low-temperature heat is made more difficult by the presence of water vapor, sulfur compounds and ammonia in the dirty gas. These substances cause the mass flow rate of the dirty gas to be higher than that for the clean gas on the cold side of the regenerator. As a result, the temperature drop on the hot side of the regenerator is significantly less than on the cold side.

One of the undesirable aspects of a low-temperature cleanup system is that the low-temperature causes most of the water vapor in the dirty gas stream to condense. This will result in the need for costly heat exchange equipment to withstand the weak acids which will be present. However, since it is necessary to cool the gas, it appears desireable to utilize as much of the heat in the dirty gas stream as is possible to improve system performance.

In addition to showing system efficiency, Figure 8-2 shows the effect of regeneration temperature on heat exchange area. While the relationship is also a function of system configuration and subject to further optimization, in general higher temperatures will require larger exchangers made out of more expensive materials. As an example, using the area requirements of Figure 8-2, a cost comparison was made for a regenerator having a cold-side outlet temperature of 750 F and one with a 1000 F outlet temperature. The results in Table 8-1 show that fabricated equipment costs were estimated to be \$17.50/ft^2 for the low-temperature unit and \$27.50/ft^2 for the 1000 F unit. In the high-temperature case, 18-8 stainless steel and chrome alloy steel are employed for the tube and shell materials. While low alloy carbon steel can be used for the shell in the low-temperature case, stainless steel would still be required for the tube service due to the H_2-H_2S environment at the operating temperature.

FIG. 8–2

EFFECT OF FUEL TEMPERATURE

BCR/SELOXOL SYSTEM

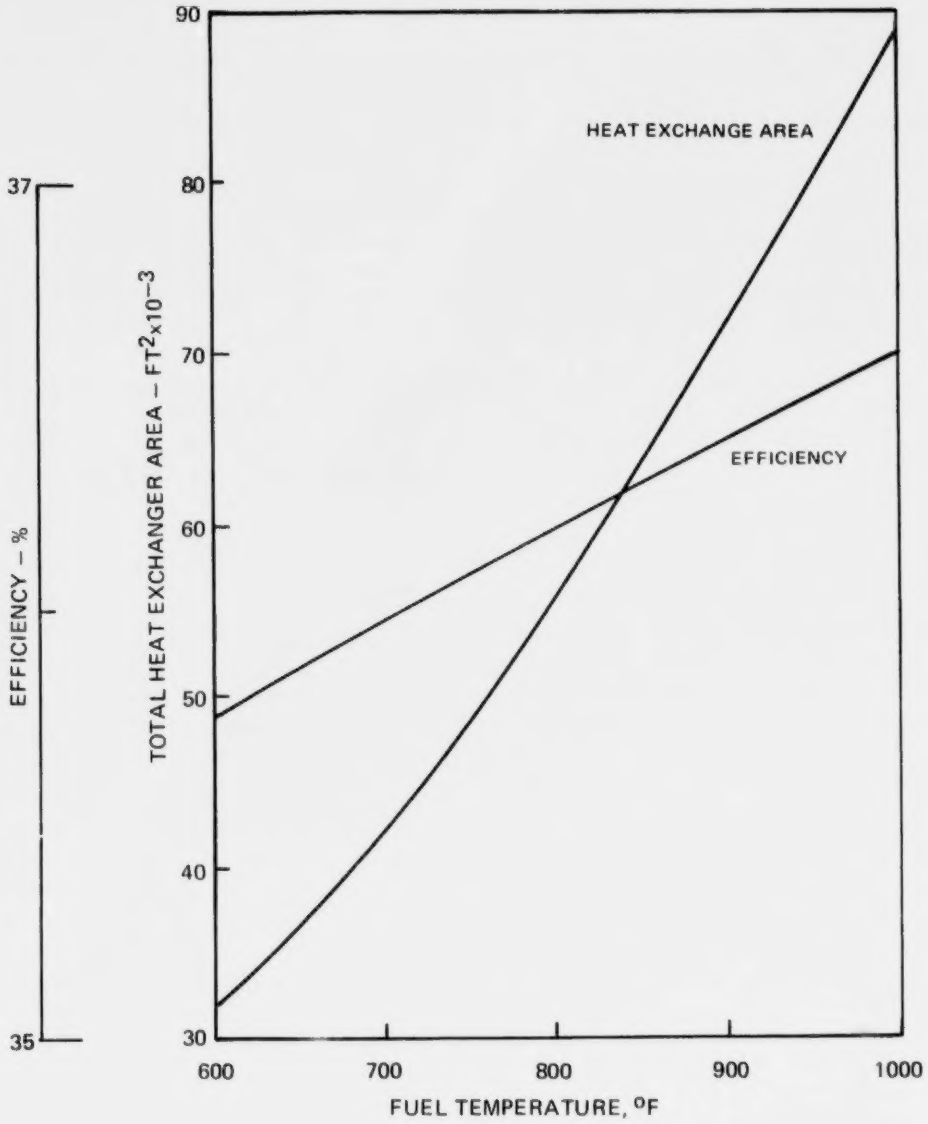

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TABLE 8-1

BCR-SELEXOL FUEL GAS REGENERATOR COSTS

Fuel Supply Temp - F	1000	750
Required Area - ft^2	88,500	48,500
Exchanger Cost - \$	3.04×10^6	1.06×10^6
Total D&E Cost - \$	5.5×10^6	1.9×10^6
Cost with interest and escalation - \$	7×10^6	2.5×10^6

The resultant delivered and erected cost differential is still quite substantial due to the significantly larger area required for the high-temperature regeneration. The incremental capital investment is approximately \$4.5/kW for a performance improvement of approximately 1.5%. This is equivalent to a decrease in heat rate of more than 120 Btu/kWhr.

Figure 8-3 presents the capital cost equivalent of heat rate improvements. As coal costs increase to more than $\$1/10^6$ Btu, an increment in capital investment of approximately \$5/kW is warranted. In the current study, a somewhat lower performance exchanger has been assumed, with the cold fuel gas being reheated to approximately 900-950 F.

SELEXOL OPERATION

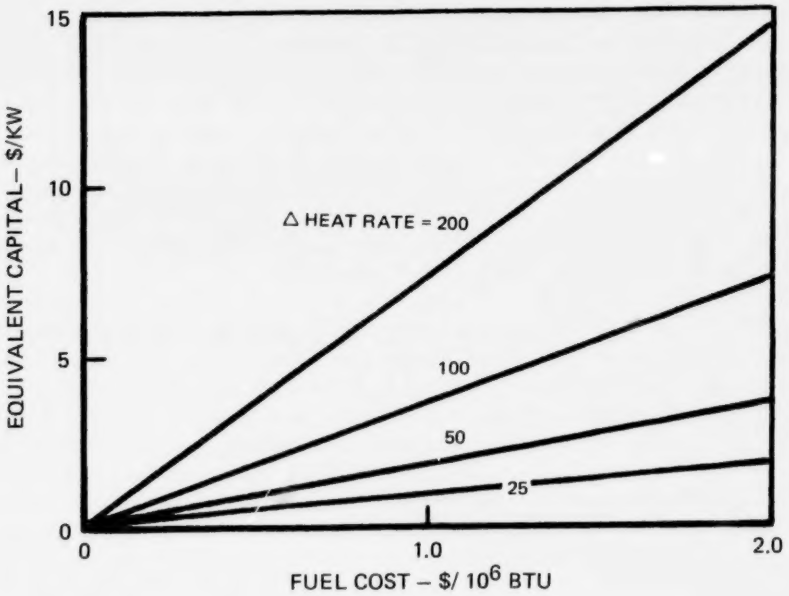
A general description of the Selexol process was given in Section 6. The basic flow diagram for the process was shown in Figure 6-7. It is a physical solvent system with absorption taking place at high pressure and stripping at ambient pressure. Refrigeration may be added to enhance solvent capacity and thereby decrease solvent flow rate, steam requirements and equipment size. A solvent flash tank with gas recycle compressor may be needed to improve selectivity. The desirability of this additional equipment is dependent upon the particular application.

One of the goals of the current study was to compare operation of the ambient-temperature Selexol process to that of the refrigerated process. Other goals initially made were the identification of the sensitivity of the ambient-temperature system to sulfur removal level and a reassessment of the advantages of catalytic conversion of COS to H_2S upstream of the Selexol absorber. The latter goal was later discarded mainly on the basis of a lack of a well defined need for such conversion. Latest estimates of COS level in the raw gas show that a system designed for H_2S removal only can still produce emissions on the order of 0.4 lb $SO_2/10^6$ Btu, well below current power plant standards. Nonetheless, technical problems are associated with CDS conversion because it is catalyzed and operates at approximately 600 to 700 F.

FIG. 8-3

CAPITAL EQUIVALENT OF HEAT RATE IMPROVEMENT

LOAD FACTOR = 0.7
ANNUAL CAPITAL CHARGE = 17%



At that point in the process there is still a significant amount of particulate matter in the stream. The problem of plugging the catalyst bed possibly could be solved by the use of a fluid bed. However, applicable experience with the required type of catalyst in this service is not available.

Refrigerated Operation

Initial Selexol sizing studies (Reference 8-6) had been based on refrigerated operation and required two basic designs, the choice of which depended upon the feed gas composition. These compositions are typified by a BuMines-type feed that has little COS and a BCR-type feed having a high level of COS. To achieve a level of 100 ppm sulfur in the clean gas requires a design based on H_2S removal in the case of the BuMines-type feed and one based on COS removal for the BCR-type feed. The feed, product, and Claus feed gas compositions are presented in Table 8-2.

For the H_2S-based design, solvent flow rate is adjusted to remove the major portion of the H_2S and is therefore capable of removing only some 30 percent of the COS. In the COS-based design, solvent flow rate is increased to remove most of the COS as well as the H_2S. Because of the relatively low solubility of COS in the solvent, a system sized for COS removal must have a proportionately higher solvent flow rate and utility consumption will increase accordingly. Moreover, the solubility of COS is only about twice that of CO_2. Simply increasing solvent flow rate would result in removal of a significant amount of the CO_2 from the fuel gas, thereby lowering its mass flow and decreasing the H_2S fraction in the Claus gas. To avoid this, the flash tank and recycle compressor shown in Figure 6-7 are included. These improve selectivity to a point where only 30 percent of the CO_2 is removed while virtually all the COS is absorbed.

A comparison of the utilities and costs associated with the two designs is given in Table 8-3. While not shown in Table 8-3, the approximate cost and utilities were estimated (Reference 8-1) for the H_2S-based design should the required sulfur removal level be relaxed to allow 1000 ppm sulfur in the clean gas. Solvent flow rate and corresponding utilities and cost are reduced by only 12 percent over the 100 ppm design. Thus, for refrigerated operation, the sensitivity to sulfur removal level appears to be quite low, at least between 100 and 1000 ppm total sulfur (actually 32 ppm and 900 ppm H_2S). Additional studies showed that for refrigerated operation, solvent flow rate and associated utilities and cost were quite insensitive to the level of H_2S in the feed within the limits studied. That is, the controlling factor in the design is that area of the absorber where the lowest sulfur levels are achieved. Thus, the Selexol system utilities and cost are largely a function of only the molar flow rate, the total pressure and, of course, the type of sulfur compound that must be removed.

Ambient Temperature Operation

Selexol designs using an ambient temperature (100 F) absorber were developed (Reference 8-2) for the same feed compositions as previously described for the regrigerated systems. Mol balances for the resulting

TABLE 8-2

BASIC SELEXOL DESIGNS

Refrigerated Operation
Removal to 100 ppm Sulfur

BuMines Type Feed (H_2S Based Design)
MOL BALANCE MOLS/HR.

	Feed	Product	Claus Gas
N_2	57,792.7	57,764.5	128.2
CO	26,616.0	26,527.6	88.4
CO_2	6,717.9	5,647.1	1,070.8
H_2	17,794.3	17,767.3	27.0
CH_4	3,486.8	3,455.6	31.2
H_2S	639.2	3.6 (32ppm)	635.6 (31%)
COS	11.6	7.6 (68ppm)	4.0
NH_3	93.0	30.5	62.5
Total	113,251.5	111,203.8	2,047.7

BCR Type Feed (COS Based Design)

	Feed	Product	Claus Gas
N_2	65,678	65,676	2
CO	26,217	26,215	2
CO_2	11,663	8,015	3,648
H_2	18,369	18,369	---
CH_4	5,202	5,198	4
H_2S	616.6	1.2 (10ppm)	615.4 (14%)
COS	141.3	10.8 (90ppm)	130.5
NH_3	64.2	1.5	62.7
Total	127,951.1	123,486.5	4,464.6

TABLE 8-3

SELEXOL UTILITIES AND COST SUMMARY
REFRIGERATED OPERATION

	Bu Mines Feed	BCR Feed
Total sulfur in product - ppm	100	100
Controlling Species	H_2S	COS
Steam (50 psi) lb/hv	105,000	305,000
Net Power - bhp	21,000	47,000
Net Power - kW	17,400	38,950
Capital Cost - $\$10^6$	21.3	49.2

systems are presented in Table 8-4. The table also gives the performance estimates for a BuMines-type feed with removal down to 500 ppm and a modified BCR-oxygen-blown feed with removal down to 60 ppm total sulfur.

Of interest is the degree of COS removal achieved at the various overall removal levels. Comparing the two BuMines cases, at 100 ppm total sulfur 46 percent of the COS is removed while at 500 ppm sulfur only 35 percent of the COS is removed. This indicates that solvent flow rate was significantly increased and that the ambient temperature design is quite sensitive to removal to levels of less than 500 ppm in the product gas.

The sensitivity of the ambient-temperature systems to sulfur removal level is shown in Table 8-5 which gives a summary of utilities and costs for each of the four designs. Comparing the two BuMines cases, decreases of 42 percent in steam consumption, 22 percent in power and 23 percent in cost are noted as the sulfur removal requirements are relaxed from 100 to 500 ppm. The COS-based design requires a flash tank and recycle compressor in addition to increased solvent flow to achieve an 18 percent H_2S level in the Claus feed gas.

Selection of Operating Temperature

It had been anticipated that the change to ambient temperature operation of the Selexol absorber would prove to be desirable both from a performance and cost standpoint. However, a review of the results shows that this is apparently not true. Tables 8-6 and 8-7 present a summary of the effect of temperature change for a BuMines-type feed where H_2S removal determines the design and a BCR-type feed where COS is the controlling species. In general, the refrigerated system requires less steam, more power, and is less costly than its ambient-temperature counterpart. The BuMines design is of particular interest

TABLE 8-4

MOL BALANCE FOR AMBIENT TEMPERATURE SELEXOL DESIGN

Bu Mines Type Feed
100 ppm Sulfur Out

	Feed	Product	Claus Gas
N_2	57,892.7	57,603.2	289.5
CO	26,616.0	26,417.3	198.7
CO_2	6,717.9	5,318.1	1,399.8
H_2	17,794.3	17,732.6	61.7
CH_4	3,486.8	3,422.4	64.4
H_2S	639.2	3.9 (35ppm)	63.3 (22%)
COS	11.6	6.3 (57ppm)	5.3
NH_3	93.0	12.6	80.4
H_2O	239.4	139.5	99.9
TOTAL	113,490.9	110,655.9	2,835.0

BCR Type Feed
100 ppm Sulfur Out

	Feed	Product	Claus Gas
N_2	65,678	65,671	7
CO	26,217	26,211	6
CO_2	11,663	9,300	2363
H_2	18,369	18,368	1
CH_4	5,202	5,194	4
H_2S	616.2	1.2 (10ppm)	615.4 (18%)
COS	141.3	11.2 (90ppm)	130.1
NH_3	64.2	---	1.5
H_2O	283	159	306
	128,234.1	124,915.4	3,434.0

Bu Mines Type Feed
500 ppm Sulfur Out

	Feed	Product	Claus Gas
N_2	57,892.7	57,666.5	226.2
CO	26,616.0	26,460.8	155.2
CO_2	6,717.9	5,624.3	1,093.6
H_2	17,794.3	17,746.1	48.2
CH_4	3,486.8	3,436.5	50.3
H_2S	639.2	46.6 (419ppm)	592.6 (25%)
COS	11.6	7.5 (67ppm)	4.1
NH_3	93.0	30.2	62.8
H_2O	239.4	135.3	104.1
TOTAL	113,490.9	111,153.8	2,337.1

BCR-Oxygen Blown Feed
60 ppm Sulfur Out

	Feed	Product	Claus Gas
N_2	381.6	379.5	2.1
CO	30,018.9	29,766.9	252.0
CO_2	10,630.5	8,113.2	2,517.3
H_2	25,070.7	24,974.1	96.6
CH_4	5,130.0	5,020.8	109.2
H_2S	954.0	0.7 (10ppm)	953.4 (22%)
COS	7.2	3.6 (53ppm)	3.6
NH_3	---	---	---
H_2O	257.4	118.2	384.3
	72,450.3	68,376.9	4,318.5

TABLE 8-5

SELEXOL COST AND UTILITIES SUMMARY

	Ambient Temperature Absorber			
	Bu Mines Feed	BCR Feed	Bu Mines Feed	O_2-Blown BCR Feed
Total S in effluent - ppm	92	100	486	63
Controlling Species	H_2S	COS	H_2S	H_2S
Steam (50 psi) - lb/hr	252,000	760,000	147,000	82,200
Net Power - bhp	5,154	30,815	4,029	1,230
Net Power - kW	4,270	25,532	3,338	1,019
Capital Cost - $\$10^6$	34.6	95.8	26.6	27.1

TABLE 8-6

REFRIGERATED VS. AMBIENT-TEMPERATURE SELEXOL OPERATION

BuMines Type Feed (H_2S Controls)
100 ppm Sulfur in Product

	Ambient-Temperature	Refrigerated	Δ
Steam - lb/hr	252,000	106,200	146,000
Net Power - kW	4,270	17,400	- 13,000
Cost - $\$10^6$	34.6	21.3	13.3

TABLE 8-7

REFRIGERATED VS. AMBIENT-TEMPERATURE SELEXOL OPERATION

BCR Type Feed (COS Controls)
100 ppm Sulfur In Product

	Ambient-Temperature	Refrigerated	Δ
Steam - lb/hr	760,000	305,000	455,000
Net Power - kW	25,530	38,940	- 13,400
Cost - $\$10^6$	95.8	49.2	46.6

in this study since that will be the basis for the integrated system designs described in Section 11. For fuel gas from the BuMines-type gasifier, refrigerated operation shows less advantage than it does in the case of the BCR-type gasifier.

The effects of changing from refrigerated to ambient-temperature operation are summarized in Table 8-8. Low-pressure steam use is debited at the rate of 20 lb/kWh while the incremental costs of the turbogenerator and condenser were \$500/kW for the low-pressure steam (exclusive of boiler). The net effect of ambient-temperature operation is a \$10/kW increase in cost and a 50 Btu/kWh decrease in heat rate. It can be seen from Figure 8-3 that this would pay off only when coal reaches a cost of approximately \$6 per 10^6 Btu. Thus, while the overall effect of a change to ambient-temperature operation would not be very significant, it is definitely not economical and would result in operation at conditions where the Selexol design is quite sensitive to sulfur removal level. Under those conditions, any unexpected factors that might harm Selexol performance would have a much greater effect on the ambient-temperature system. Therefore, refrigerated operation was chosen as the design basis for this study.

Effect of Sulfur Removal Level

Based on the data presented in Tables 4-3 through 4-5, it is clear that the major factor determining Selexol system utility consumption and cost is whether the system design is based on H_2S or COS removal. For the H_2S-based design with refrigerated operation, only a 12% decrease in solvent flow rate (and cost) was noted when the product gas was allowed to increase from 100 to 1000 ppm total sulfur (32 to 900 ppm H_2S). Thus, it appears that at refrigerated operating conditions, there is only a slight penalty involved with removal to low H_2S levels. However, in the case of ambient operation with the same feed composition, a 30 percent increase in cost and 60 percent increase in steam consumption is noted as removal level is reduced from 500 to 100 ppm total sulfur (420 to 35 ppm H_2S). Thus, it appears that at ambient temperature the system is somewhat more sensitive to the degree of H_2S removal.

STEAM TO COAL FEED RATIO

Earlier studies (e.g., Reference 8-3) have shown the benefits to be achieved by lower steam/carbon feed ratios. With coal as the feed, this is usually referred to as the steam/coal ratio. For combined-cycle power generation, methane and hydrogen yields are not important so long as the resultant gas will permit stable combustion. This allows a great degree of latitude in operating conditions and generally would permit gasification with no hydrogen other than that contained in the coal.

To test the effect of steam/carbon feed ratios, it is of interest to review the somewhat easier case of an oil gasifier using the equilibrium calculation procedure for the gasifier. Typical ranges for steam/oil ratios would be 0.0 to 0.4 (lb/lb) while the air/oil ratios would be 6.0 to 6.5. The minimum air/oil ratio that would provide 1 atom of oxygen per atom of

TABLE 8-8

SUMMARY OF PERFORMANCE AND COST
REFRIGERATED VS. AMBIENT TEMPERATURE SELEXOL OPERATION

Performance Effect of Change from Refrigerated to Ambient Temperature

	Change	Performance Effect
Stream Consumption	+146,000 lb/hr	- 7,300 kW
Power Consumption	- 13,000 kW	+13,000 kW
Net Effect on Output		+ 5,700 kW
Net Effect on Heat Rate		- 45 Btu/kWh

Cost Effect of Change from Refrigerated To Ambient Temperature

Change in Selexol Cost	+ $13.3 x 10^6
Change in LP Steam Turbine and Condensor	- 3.6 x 10^6
Net Capital Cost Change	+ 9.7 x 10^6
Net Change in Cost Per RW	+ 9.6/kW

carbon is 5.0. However, to achieve a reasonable reactor size, it is necessary to increase this to 6.0 thereby providing sufficient air to increase combustion temperature and, thereby, reaction rates.

Table 8-9 shows that change in volumetric heating value of the fuel gas is very small over the range of operational steam/oil ratios. When viewed in terms of chemical heating value per pound of oil consumed, the output of the gasifier is constant over the range of steam/oil ratios considered. Also shown in Table 8-9, the effects of steam addition on composition are an increase in hydrogen and carbon dioxide production coupled with a decrease in CO. Stoichiometrically, each additional molecule of hydrogen in steam brings with it an oxygen atom which will react with one CO molecule to form CO_2. Each added H_2 molecule means one less CO molecule. Moreover, the higher heating value of H_2 and CO are almost the same so their effects on gas heating value cancel. Further examination of the product gas shows that about 25 percent of the input steam shows up as hydrogen and the remainder leaves as water vapor in the product gas. Thus, the net effect of steam addition on the product gas is minimal and the heat needed to raise the steam is mostly lost since the latent heat cannot practically be recovered from the water vapor in the fuel gas.

To maintain reactor temperature, air/oil ratio would have to be increased, thereby reducing chemical heating value. Even if it were possible to keep the water vapor in the fuel gas stream (as with hot cleanup) it would be used at a relatively low efficiency since gas turbine expansion ratios are significantly less than those of steam turbines.

In the case of low-temperature cleanup systems, virtually all water vapor is condensed during the gas cooling process. One method of recovering some of this heat is by using the latent heat to heat water from a fuel gas resaturator (Figure 8-4). This increases the mass of the fuel gas stream. The effect on performance is illustrated by studies of resaturation of the clean gas in the BuMines/Selexol system.

The potential performance improvement to be achieved with the addition of water vapor to the fuel gas was estimated by varying the cleanup system output composition. The results are shown in Figure 8-5. In determining performance, no penalty was associated with the addition of the vapor so the trends shown in that curve are the maximum that can be achieved. In practice, heat must be provided for humidification and could detract from performance. In terms of water used per megawatt of power, the incremental power produced requires about 14,000 lb/hr for each additional megawatt of electrical power. For this system a 300 F dew point will produce a water vapor mol fraction of .248. This was taken as a practical maximum since further resaturation would require the use of heat at temperatures over 300 F while producing power at a relatively high heat rate of 15000 Btu/kWh.

AIR-BLOWN BCR GASIFIERS

The earlier phases of this study had used available data for the BCR-type gasifier that had a high steam feed rate and was more suited to synthesis gas

TABLE 8-9

EFFECT OF STEAM ADDITION ON FUEL GAS CHEMICAL HEATING VALUE

Fuel - Venezuelan Residual Oil

Air/Oil Ratio = 6.0

	Steam/Oil Ratio		
Fuel Gas Characteristics	0.	0.2	0.4
Mole Fraction H_2	.1469	.1514	.1548
Mole Fraction H_2O	.0335	.0612	.0858
Mole Fraction CO	.2335	.215	.1985
Mole Fraction CO_2	.0154	.0248	.0328
H_2 SCF/lb Oil	16.66	17.83	18.91
CO SCF/lb Oil	26.49	25.32	24.24
HHV Btu/SCF	122.9	118.4	114.2
Gas Produced SCF/lb Oil	113.43	117.78	122.13
Output Gas HHV - Btu/lb Oil	13940	13945	13947

BUMINES/SELEXOL SYSTEM WITH FUEL GAS RESATURATION

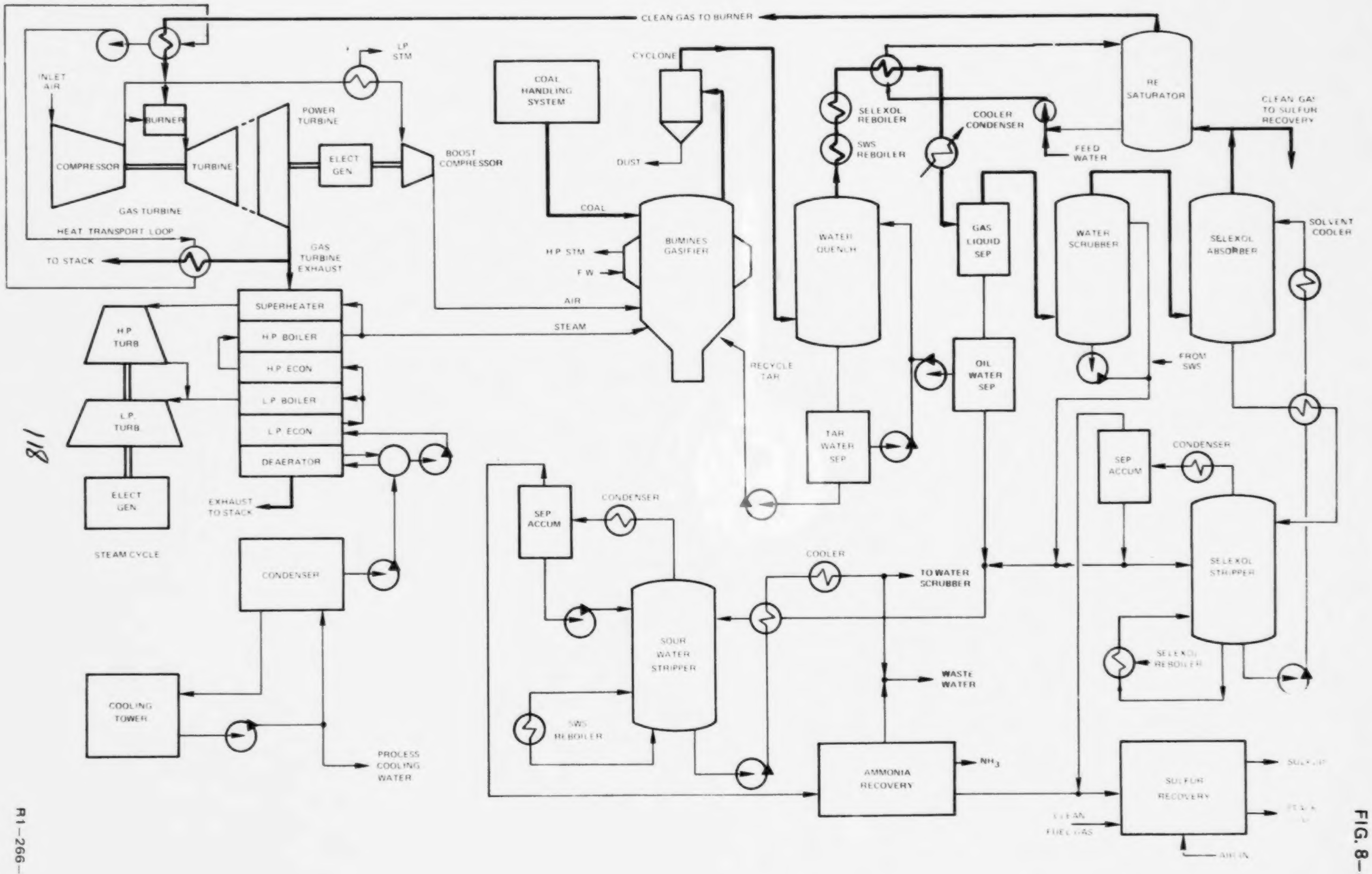

CLEAN GAS TO BURNER
INLET AIR
BURNER
COMPRESSOR
TURBINE
POWER TURBINE
GAS TURBINE
HEAT TRANSPORT LOOP
TO STACK
LP STM
ELECT GEN
BOOST COMPRESSOR
GAS TURBINE EXHAUST
SUPERHEATER
H.P. BOILER
H.P. ECON
L.P. BOILER
L.P. ECON
DEAERATOR
H.P. TURB
L.P. TURB
ELECT GEN
EXHAUST TO STACK
STEAM CYCLE
CONDENSER
COOLING TOWER
PROCESS COOLING WATER
COAL HANDLING SYSTEM
COAL
CYCLONE
DUST
BUMINES GASIFIER
H.P. STM
F.W.
AIR
STEAM
RECYCLE TAR
WATER QUENCH
TAR WATER SEP
SELEXOL REBOILER
SWS REBOILER
COOLER CONDENSER
GAS LIQUID SEP
OIL WATER SEP
RE SATURATOR
FEED WATER
WATER SCRUBBER
FROM SWS
SELEXOL ABSORBER
SOLVENT COOLER
CLEAN GAS TO SULFUR RECOVERY
SEP ACCUM
CONDENSER
SELEXOL STRIPPER
SELEXOL REBOILER
SEP ACCUM
CONDENSER
SOUR WATER STRIPPER
SWS REBOILER
COOLER
TO WATER SCRUBBER
WASTE WATER
AMMONIA RECOVERY
NH_3
SULFUR RECOVERY
SULFUR
CLEAN FUEL GAS
AIR IN

FIG. 8–4

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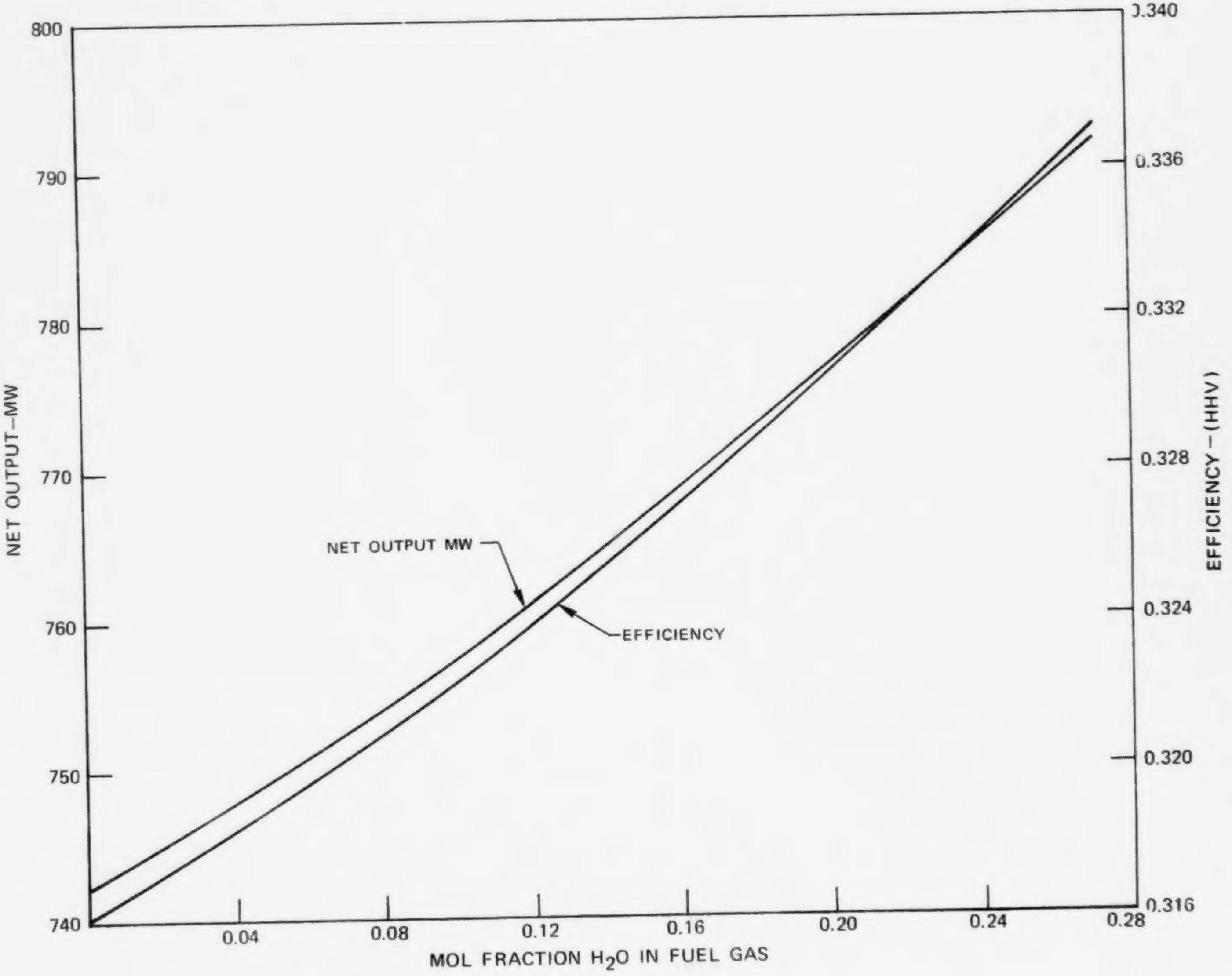
FIG. 8–5

EFFECT OF WATER VAPOR IN FUEL GAS BUMINES–SELEXOL SYSTEM



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or methane production. Work subsequently performed under EPRI-sponsorship (Reference 8-4) has resulted in descriptions of gasifier operation at lower steam feed rates that are more suited to combined-cycle power generation. A comparison of the two estimates is given in Table 8-10. The improved cold gas efficiency is a good indication of relative performance but does not account for all of the factors that will affect system efficiency.

A careful inspection of the data in Table 8-10 shows that the remarkable increase in cold gas efficiency of nearly six percentage points is due only in part to the reduced steam feed rate. The reduction in transport gas requirements is also quite significant. Transport gas is compressed to higher than gasifier pressure and must be cooled prior to compression. A considerable amount of heat must be supplied via combustion in the gasifier to increase both transport gas and steam temperatures. Surprisingly, while the absolute hydrogen yield decreased with reduced steam, the mol fraction of hydrogen increased. Thus, there is certainly no reason to expect a combustion problem as a result of the change.

Low Temperature Desulfurization

In terms of sulfur removal, a more desirable ratio of H_2S to COS is evident. While total sulfur molar flow rate remains the same, the higher partial pressure resulting from lower total molar flow rate will mean a corresponding reduction in Selexol system size, cost, and utilities.

The reduced CO_2 concentration could permit the use of a nonselective desulfurization process. The effect of substituting a nonselective cleanup process for the Selexol system was evaluated using utilities typical of a hot-potassium system. The effect on overall performance was estimated to be negligible. However, the Selexol process was retained as the baseline system for the study. It gives reasonable performance representative of low-temperature processes and will probably result in cost estimates that are slightly conservative.

High-Temperature Desulfurization

The Conoco half-calcined dolomite absorber is quite sensitive to both temperature and gas composition. Reactions in the absorber closely approach equilibrium (Reference 8-5) and desulfurizer performance can be estimated using data for H_2S and COS reactions with the dolomite. H_2S absorption involves the following reaction:

$$[CaCO_3 \,.\, MgO] + H_2S \qquad [CaS \,.\, MgO] + H_2O + CO_2 \tag{8-3}$$

for which the equilibrium H_2S concentration in the gas is given by

$$[H_2S] = [H_2O]\,[CO_2]\;P/K \tag{8-4}$$

TABLE 8-10

COMPARISON OF BCR DATA

Coal - Illinois No. 6 - 700, 000 lb/hr.

	Air-Blown Gasifier			
	1975 EPA Study		EPRI Data	
	Mols/Hr (1)	Mol %	Mols/Hr (1)	Mol %
CH_4	5,188	3.65	3,775	3.37
H_2	18,319	12.88	15,315	13.68
CO	26,151	18.38	32,190	28.75
CO_2	11,751	8.26	3,396	3.03
H_2S	687	0.48	751	0.67
COS	142	0.10	76	0.07
N_2	65,499	46.04	53,753	48.02
NH_3	573	0.4	479	0.43
H_2O	13,953	9.81	2,213	1.98
	142,263		111,948	
HHV-Btu/SCF	138.0		171.2	
Air/Coal Ratio	3.09		2.78	
Steam/Coal Ratio	.567		.144	
Transport Gas/ Coal Ratio	.426		.088	
Cold Gas Eff. (2)	78.5%		83%	

(1) Includes transport gas

(2) Based on net gas make - excluding transport gas

where [ ] = mol fraction of component

P = total pressure, atm

K = equilibrium constant

From Figure 8-6 and Equation 8-4, it is clear that the degree of H_2S removal increases with temperature for a given gas composition. Also, the degree of removal at a given temperature is inversely proportional to temperature and concentration of the reaction products, CO_2 and H_2O.

Although no data have been reported for COS adsorption by half-calcined dolomite, high COS removal efficiencies are predicted thermodynamically according to the reaction:

$$[CaCO_3 \;.\; MgO] + COS \qquad [CaS \;.\; MgO] + CO_2 \tag{8-5}$$

where the residual COS concentration is

$$[COS] = [CO_2]^2 \; P/K \tag{8-6}$$

The equilibrium constant for this reaction is shown in Figure 8-7. The degree of COS removal like that for H_2S increases with increasing temperature and decreasing CO_2 concentration.

From the above, it would appear that reduced steam feed to the gasifier with its associated reduction in H_2O and CO_2 in the off-gas should improve desulfurization performance. However, with the low H_2O and CO_2 content, a third reaction becomes important, that of decomposition of the acceptor. It imposes a maximum operating temperature for this process depending on the partial pressure of carbon dioxide in the gas phase; i.e., the temperature should not exceed that at which the CO_2 partial pressure is equal to the decomposition pressure for $CaCO_3$ via the following endothermic reaction:

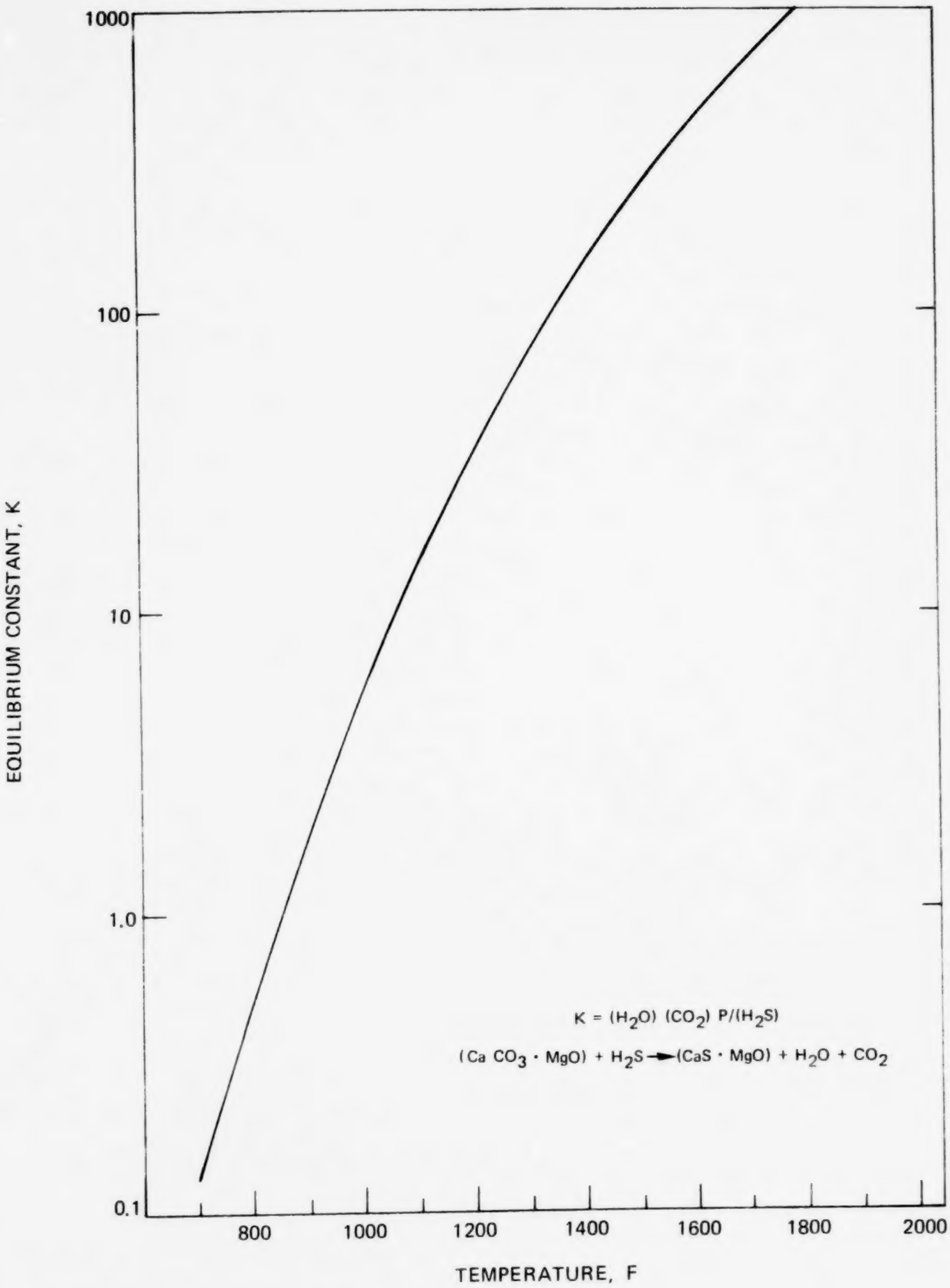
$$Ca\,CO_3 \qquad CaO + CO_2 \tag{8-7}$$

The equilibrium decomposition pressures for calcium carbonate are given in Figure 8-8.

An example of the characteristics of the Conoco absorption process is given in Figure 8-9 using a BCR-type feed having a high water vapor and CO_2 composition; the partial pressure of CO_2 in the product gas is shown as a solid line for each of two different operating pressures (the variation with temperature is due to the water gas shift reaction). The limiting or minimum partial pressure shown by the dashed line is the equilibrium partial pressure of CO_2 above $CaCO_3$. The partial pressure of CO_2 must be greater than this to prevent decomposition of the $CaCO_3$ to CaO. The intersection (Point A) indicates that at the absorber operating pressure of 450 psia, the maximum operating temperature could be 1780 F and the residual sulfur would be 325 ppm

FIG. 8 -6

EQUILIBRIUM CONSTANT FOR H_2S ABSORPTION BY HALF – CALCINED DOLOMITE



$K = (H_2O)\,(CO_2)\,P/(H_2S)$

$(CaCO_3 \cdot MgO) + H_2S \rightarrow (CaS \cdot MgO) + H_2O + CO_2$

FIG. 8–7

EQUILIBRIUM CONSTANT FOR COS ABSORPTION BY HALF–CALCINED DOLOMITE

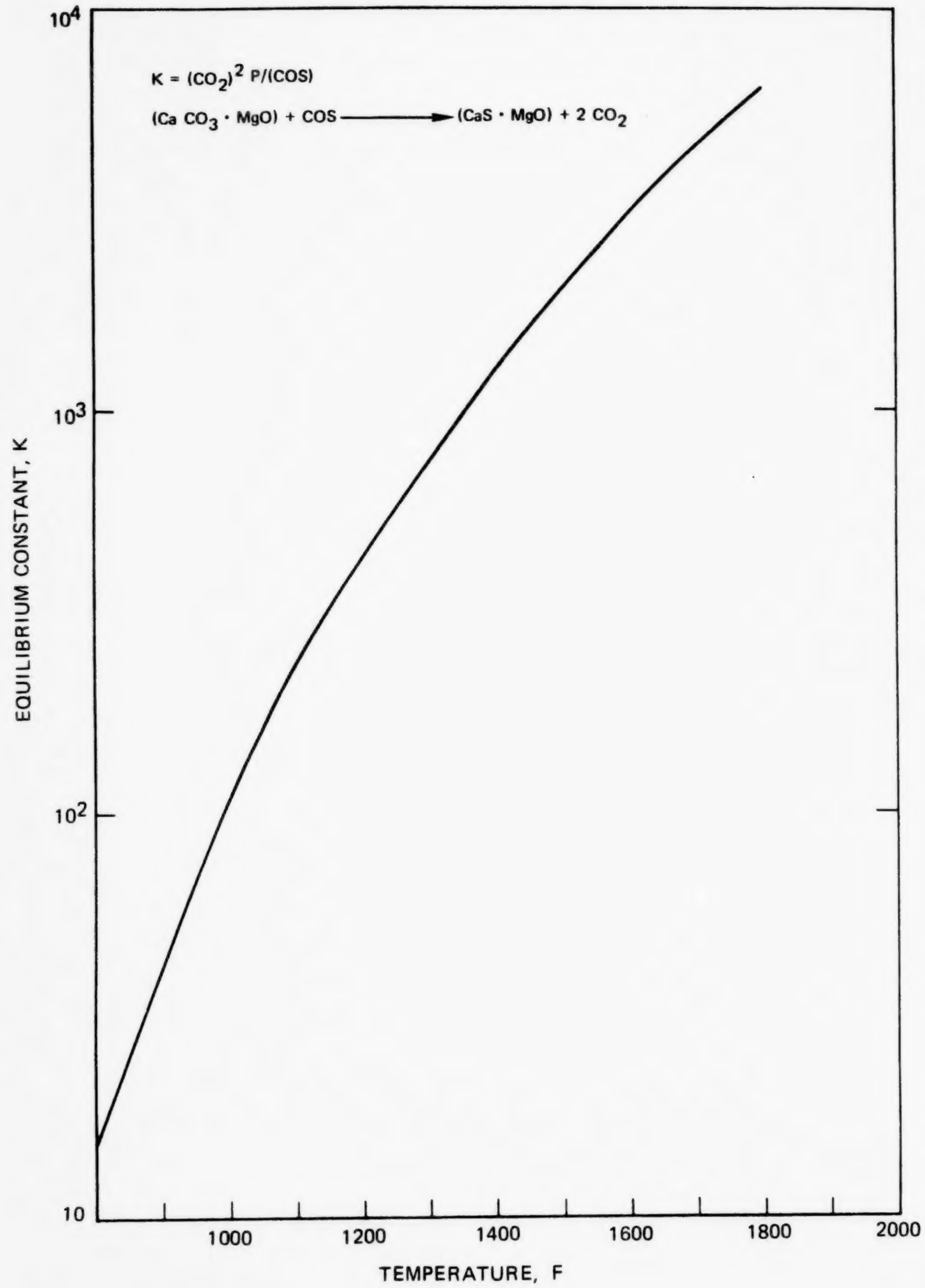


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FIG. 8–8

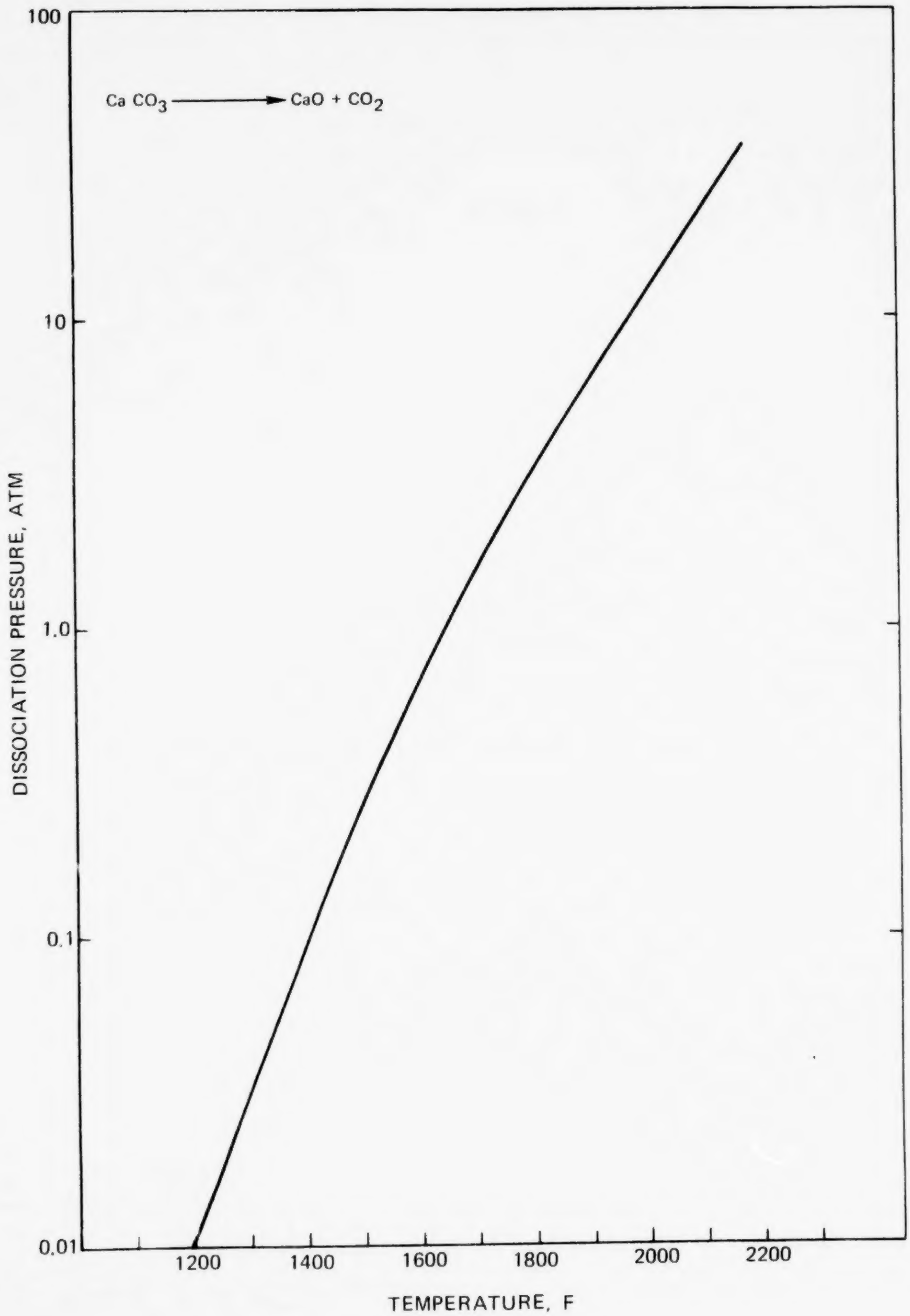

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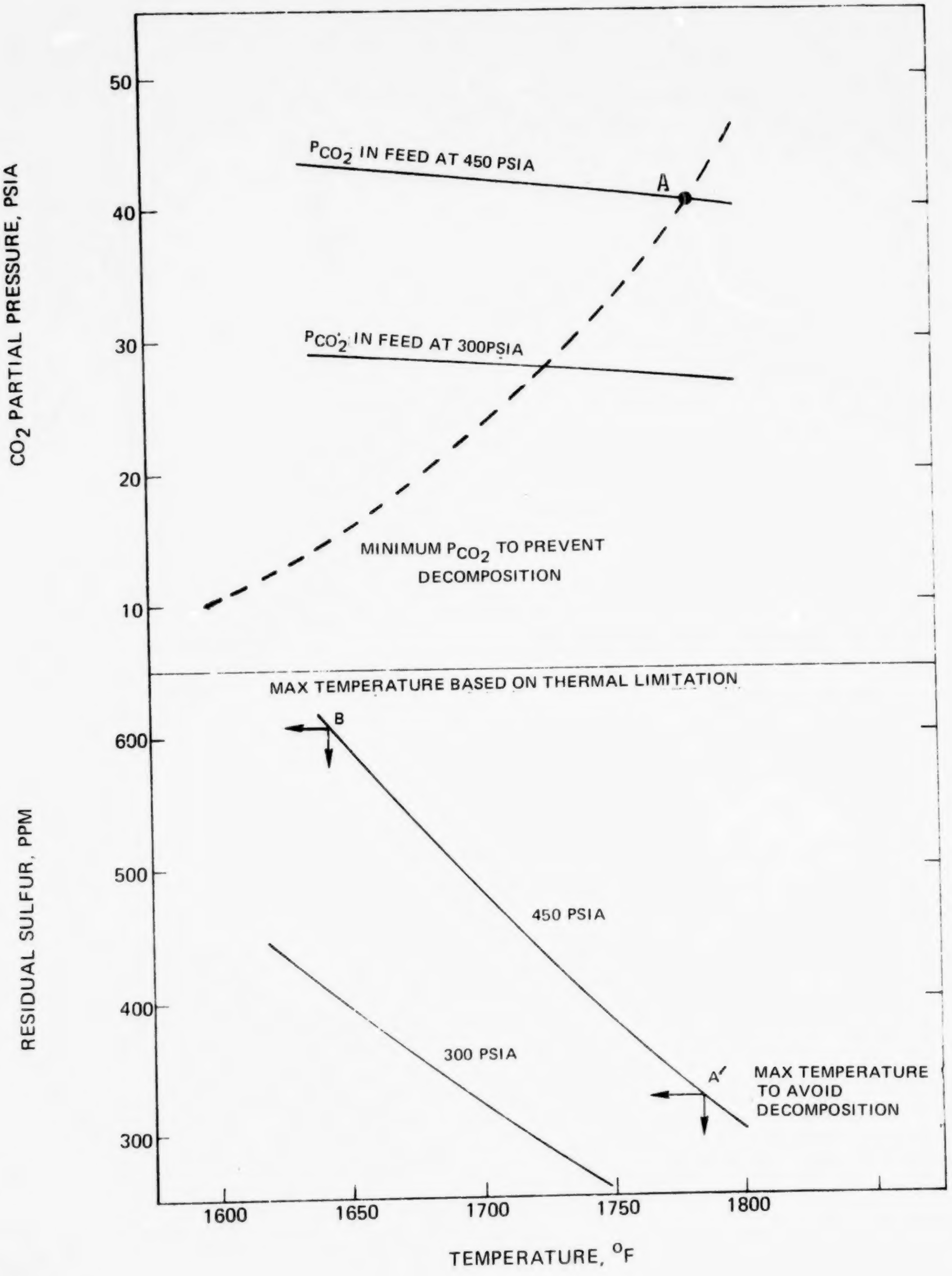
FIG. 8–9

SULFUR ABSORPTION BY HALF CALCINED DOLOMITE



(Point A'). However, the product gas enters the absorber at only 1750 F and heat balance constraints limit the maximum temperature to about 1650 F since the combined chemical reactions are endothermic to the extent of 112 Btu/mol of feed gas. The residual sulfur content at this temperature is around 600 ppm (Point B in Figure 8-9).

In the case of the air-blown BCR gasifier with low steam feed, desulfurizer pressure is 340 psia and partial pressure of carbon dioxide in the feed is only 10 psia. Thus the temperature in the absorber must be limited to 1600 F to avoid decomposition. For that reason, the raw gas which leaves the gasifier at 1700 F must be cooled prior to desulfurization. The lower temperature reduces the equilbrium constants in Equations (8-4) and (8-6), but that reduction is less than 50 percent while CO_2 and H_2O concentrations are reduced by a factor of 2.5 and 5.0 respectively. The resultant sulfur in the gas (based on equilibrium) is only 106 ppm as opposed to approximately 600 ppm for the high-steam feed case.

PARAMETRIC STUDIES OF THE U-GAS/CONOCO SYSTEM

Because of the potential benefits of reduced steam feed rates, the effect on the U-Gas system was investigated. The gasifier equilibrium model developed at UTRC was used to predict the effect of variations in steam feed rates while maintaining a fixed reaction temperature and constant heat leak. Discussions with IGT supported the validity of the study and led to their recommendation of 0.25 lb H_2O/lb coal as the minimum practical ratio for the U-Gas process.

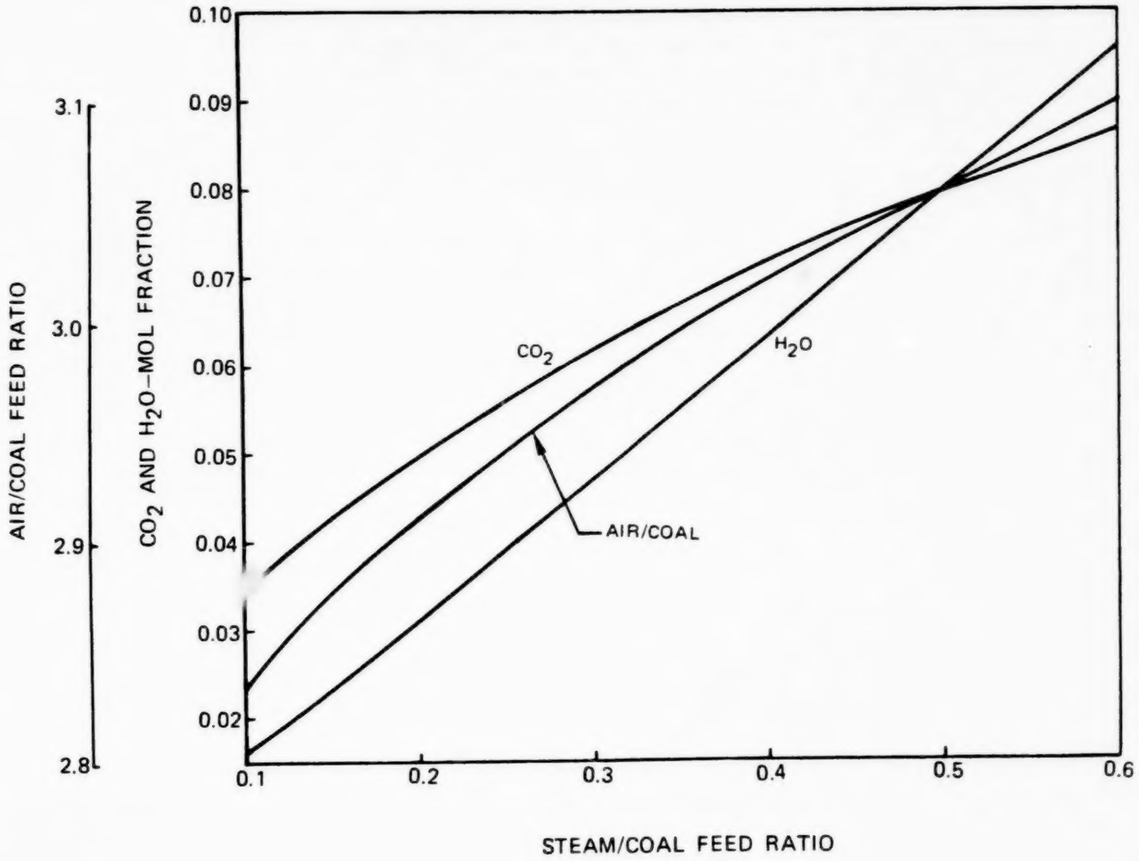
For an Illinois No. 6 coal feed, the concentration of H_2O and CO_2 in the raw gas as a function of steam/coal ratio to the gasifier is given in Figure 8-10. The effect on sulfur removal is shown in Figure 8-11. A steam/coal feed ratio of 0.1 is limiting from a theoretical standpoint in that the CO_2 partial pressure is too low to prevent decomposition of the carbonate. Also, as CO_2 concentration is reduced, there is the potential for carbon formation via the Boudouard reaction. A steam/coal ratio of 0.2 coupled with the residual moisture in the coal gives a total H_2O/ coal ratio of 0.24. This value was selected as the basis for operation of a U-Gas/Conoco high-temperature desulfurization system. The estimated product gas characteristics and high-temperature desulfurizer performance for this system are given in Table 8-11.

It should be noted that in both the BCR and U-Gas cases, the desulfurizers have been assumed to run at a temperature approximately 100 F lower than that of the Conoco design upon which performance and cost is based. Quantitative evaluation of the effect on approach to equilibrium of that difference was not available. A longer residence time will most likely be required resulting in a larger desulfurization system.

GASIFIER MODELING

Modeling techniques developed in earlier phases (References 8-3 and 8-6) were used to verify data obtained during this program. In addition to use of

U–GAS GASIFIER EQUILIBRIUM COMPOSITION AT 2010R



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HIGH–TEMPERATURE CLEANUP SYSTEM PERFORMANCE

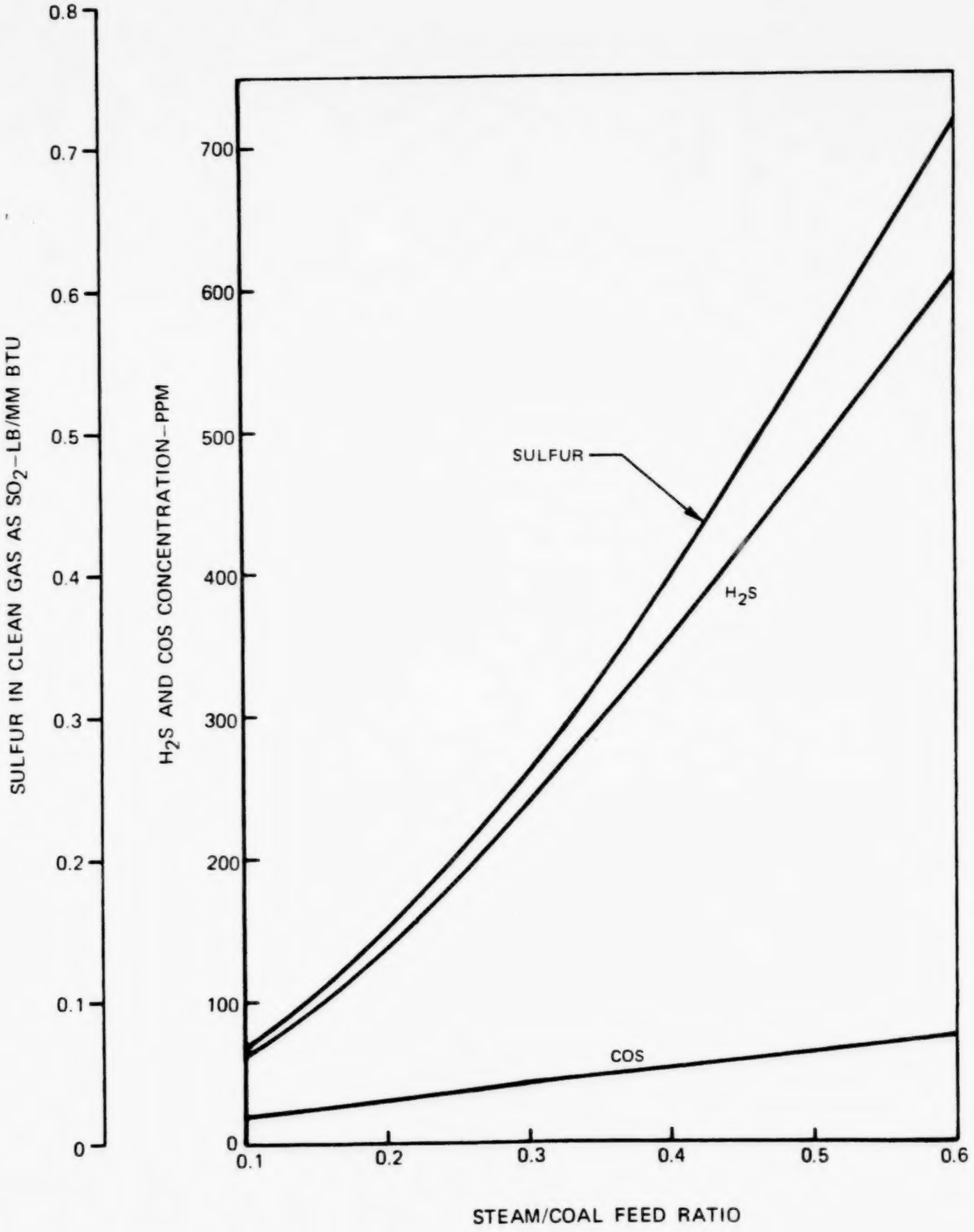

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TABLE 8-11

COMPARISON OF HIGH VS. LOW STEAM PERFORMANCE

U-Gas Gasification/Conoco Desulfurization
Coal Feed - Illinois #6 - 700,000 lb/hr

	High Steam Feed		Low Steam Feed	
Comp.	Gasifier Out Mol/Hr	Desulfurizer Out Mol/Hr	Gasifier Out Mol/Hr	Desulfurizer Out Mo/Hr
CH_4	4,002	4,002	3,724	3,724
H_2	20,335	21,498	16,926	17,003
CO	23,738	22,574	29,126	29,049
CO_2	11,178	13,071	5,832	6,732
H_2S	779	90	763	17
COS	32	12	41	3
N_2	57,801	57,801	55,053	55,053
NH_3	42	42	31	31
H_2O	14,013	13,539	3,606	4,275
Total	131,920	132,629	115,102	115,887
HHV-Btu/SCF	138.8	138.1	162	160.9
Air/Coal Ratio	3.01		2.86	
Steam/Coal Ratio	0.557		0.2	
Transport Gas/ Coal Ratio	0.053		0.053	
Cold Gas Efficiency - %	80.2		81.5	

the equilibrium model in evaluating the effects of steam/coal ratio, it was used to identify the possible pollutant compounds and potential levels that might be expected for each of the gasifiers studied. These are presented in Tables 8-12 and 8-13. While the results given in Tables 8-12 and 8-13 are subject to the limitations of an equilibrium calculation not absolute in character, they are useful to get a feel for the degree of optimism or pessimism about pollution control for the fuel gas from each type of gasifier.

TABLE 8-12

GASIFIER EQUILIBRIUM COMPOSITIONS

	BCR/Air-Blown					BCR/O_2 - Blown			U-Gas		
	1975 Data		Current Data (Low Steam Feed)								
	Predicted	Adiabatic Equilib.	Predicted	Adiabatic Equilib.	Isothermal Equilib.	Predicted	Adiabatic Equilib.	Isothermal Equilib.	Predicted	Adiabatic Equilib.	Isothermal Equilib.
H_2O	.0981	.0816	.0198	.0177	.0160	.1440	.1339	.1272	.1062	.0894	.0893
CO	.1838	.1920	.2875	.2908	.2997	.3533	.3527	.3760	.180	.1851	.2058
H_2	.1288	.1693	.1368	.1511	.1648	.2950	.3183	.3462	.1541	.1883	.2052
CO_2	.0826	.0804	.0303	.0276	.0214	.1251	.1237	.1012	.0847	.0852	.0698
N_2	.4604	.4483	.4802	.4762	.4693	.0045	.0075	.0072	.4382	.4278	.4179
CH_4	.0365	.0222	.0337	.0291	.0215	.0604	.0528	.0315	.0303	.0178	.0058
NH_3	.004	.00046	.0043	.00023	.00022	.0064	.00009	.00008	.00032	.00035	.00028
COS	.001	.0002	.00067	.00042	.0004	.0010	.00038	.00036	.00024	.00018	.00018
H_2S	.0048	.00543	.0067	.0069	.0068	.0103	.0107	.0103	.0059	.0058	.0057
COS/H_2S	.021	.037	0.1	.061	.059	0.1	.035	.035	.041	.031	.031
T-°F	1750	1595	1700	1636	1700	1700	1619	1700	1660	1533	1660

TABLE 8-13

MOLTEN SALT GASIFIER EQUILIBRIUM COMPOSITION

Mole Fractions	Kellogg Data	Equilibrium	Equilibrium
N_2	0.4969191	0.48962	0.48957
CO	0.2768238	0.29089	0.29085
H_2	0.1347269	0.15312	0.15290
CO_2	0.0301001	0.022750	0.022810
H_2O	0.0229448	0.018882	0.018908
CH_4	0.0146700	0.0068163	0.0067769
$Na_2CO_3(\ell)$	0.0113919	0.011713	0.011699
$Na_2S(\ell)$	0.0052493	0.0056839	0.0056828
C(s)	0.0041054	-	-
NaOH(g)	0.0020429	0.0000240	0.0000239
H_2S	0.0009604	0.0004741	0.0004760
COS	0.0000654	0.0000292	0.0000293
NH_3			0.0001790
HCN			0.0000560
NaCN(g)			0.0000283
Na(g)			0.0000070
HCNO			0.0000014
CH_2O			0.0000013
C_2H_4			0.0000009
$(NaCN)_2$			0.0000004
HCO			0.0000003
CH_3			0.0000001
NaH(g)			0.0000001
Totals	1.0000000	1.0000025	1.0054637

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SECTION 9

POWER SYSTEM EMISSIONS

INTRODUCTION

A significant advantage of integrated coal gasification/gas cleanup/ combined cycle power plants is the reduced amount of air pollutants they can be designed to emit as compared to conventional coal fired plants with flue gas desulfurization. This section discusses those emissions for which major control equipment has been incorporated or where special component designs are required; i.e., sulfur dioxide, nitrogen oxides and particulates. The emissions of these primary pollutants are summarized in Table 9-1 for the seven power plant configurations under consideration.herein. In all low-temperature cases, emission levels will be equal to or better than applicable standards. However, the systems with high-temperature cleanup will require additional fuel processing or combustion modification to reduce nitrogen oxides resulting from ammonia or other nitrogen compounds in the fuel.

Sulfur levels given in Table 9-1 include the effect of Claus plant tail-gas cleanup and are nearly all due to the presence of H_2S and COS in the fuel. Low-temperature designs are quite sensitive to COS and increases in the COS level could have a significant effect on desulfurizer performance or cost. High-temperature desulfurizer performance estimates are based on an equilibrium calculation and, to avoid decomposition of calcium carbornate as a result of the low CO_2 concentration in the fuel gas, are based upon operating temperatures some 100 F lower than those at which Conoco did testing.

NO_x levels in Table 9-1 assume that all ammonia is converted to NO_x. The use of a pre-mixed combustor promises to reduce thermal NO_x formation by 80 percent over conventional combustors. Test data were developed by Turbo Power and Marine Systems, a subsidiary of United Technologies Corporation, in cooperation with the Texaco Development

TABLE 9-1

EMISSION SUMMARY

		Low-Temperature Cleanup				High-Temperature Cleanup		
Emission lb/10^6Btu	U-Gas High-Steam Selexol	U-Gas Low-Steam Selexol	BCR Air-Blown Selexol	BCR O_2-Blown Selexol	Molten Salt	U-Gas Low-Steam Conoco	BCR Air-Blown Conoco	BCR O_2-Blown Conoco
Sulfur								
Gas Turbine Exhaust	0.18	0.23	0.39	0.37	0.21	0.15	0.09	0.69
Sulfur Recovery	0.002	0.003	0.006	0.006	0.004	0.04	0.04	0.04
Nitrogen Oxides								
Thermal	< 0.35	< 0.35	< 0.35	> 0.35	< 0.35	< 0.35	< 0.35	> 0.35
Fuel Bound(1)	-	-	-	-	-	0.17(2)	2.52	2.60
Particulates	.01	.01	.01	.01	.01	.03	.03	.03

---

(1) Based on 100% conversion to NO

(2) NH_3 in fuel gas based on equilibrium calculation - actual likely higher

Company. These data demonstrate this reduction when various low-, medium- and high-Btu gases are burned in a pre-mixed burner. Extrapolation from current data shows that levels of 75 ppm (corrected to 15% O_2 in the exhaust) could be achieved with the pre-mix concept in all but the oxygen-blown systems. For those medium-Btu gases, the fuel temperature may be limited to some value lower than 1000 F in order to limit NO_x formation.

There are problems associated with a pre-mix design, but ongoing research and development should solve them. Unfortunately, the pre-mix concept relies on fuel lean combustion to achieve the lower thermal NO_x emission level. Under these conditions virtually all fuel-bound nitrogen (NH_3) is converted to NO. Thus, where there is NH_3 in the fuel, an alternate approach will be needed. Tests have been run with staged combustion techniques using a fuel rich burn followed by a rapid quench. The results show effective reduction of NO production. An evaluation of combustion modification techniques as opposed to fuel processing is needed to select the best approach.

Erosion of gas turbine blading limits the quantity and size of particulate matter that may be in the fuel. Low-temperature particulate removal processes are available to meet these limits. No high-temperature process has been demonstrated to be effective enough. Such a process is essential to the use of high-temperature desulfurization, and its feasibility has been assumed herein when postulating high-temperature desulfurization systems. Since limited data are available, a definitive particulate cleanup requirement cannot be stated. It is not clear whether or not high-temperature systems will be able to meet the needs of advanced gas turbines in this important area. It is believed that current estimates are conservative and could be met by a water wash. Of course, a water wash would obviate the basic advantage of a high-temperature desulfurizer.

DEFINITION OF GOALS FOR CLEANUP SYSTEMS

Absolute guidelines are lacking for the degree of cleanup needed to avoid adverse health and ecological effects. The capabilities of the available or potentially available equipment here played an important part in arriving at the goals for the systems studied herein. Additionally, there are basic guidelines, such as current or proposed EPA standards for gas turbine and other power generating systems. They must be equalled or bettered for the systems to be attractive. These emission standards are presented in Table 9-2.

Because of the high metal temperatures, corrosion and erosion can be a problem in gas turbines. As a result, gas turbine manufacturers have established stringent specifications for fuels to be burned in industrial applications. A summary of pertinent specifications is given in Table 9-3. The major problem in current use of gas turbines by utilities is hot end corrosion. This results from alkali metal salts and sulfur in the fuel. The corrosion agent is the alkali metal sulfate which attacks the oxidation-

TABLE 9-2

COMPARISON OF EMISSION STANDARDS

	Standards For Conventional Plants		Proposed for Gas Turbines[(1)] (Ref. 9-1)
	Coal-Fired	Oil-Fired	Distillate-Fired
Sulfur as SO_2 lb/MMBtu	1.2	0.8	1.0
Nitrogen Oxides as NO_2-lb/MMBtu	0.7	0.3	0.35 to 0.6 (2)
Particulates lb/MMBtu	0.1	0.1	None

---

(1) Standards are given in ppm in stack gas with 15% oxygen.

(2) Variation is due to allowance for fuel-bound nitrogen - additional allowance made based on heat rate.

resistant coatings. Once these coatings have been penetrated, rapid oxidation of the base alloy occurs. Thus, although the turbine could withstand fuels having sulfur contents well above those allowed by environmental constraints, the presence of alkali metals reduces the allowable content to a much greater degree. The presence of particulates also contributes to corrosion since any spalling of the coatings due to impingement could result in subsequent oxidation of the base alloy.

Particulate Loading

One of the major concerns in setting fuel specifications is the particulate content. As can be seen in Table 9-3, the particulate loadings are low, varying from less than 1 ppm to about 30 ppm. The reason for the low allowable loadings is the potential erosion of the turbine blades.

A number of attempts have been made to utilize coal directly in a turbine (References 9-2, 9-3, 9-4). Due to excessive erosion, all have failed to obtain reasonable machine lifetime. Yet, the actual limitations of particulate content or size distribution have not been adequately defined. A previous UTRC study (Reference 9-5) on methods of cleaning emissions from jet engine test cells showed that the mean diameter of the particles emitted from liquid-fuel burning engines was 0.1 μ with over 99 percent of the particles being less than 1 μ. Since erosion is not a problem with these engines, it can be concluded that particles of 1 μ or smaller in the fuel or combustion products are not harmful to engines. In fact, it was pointed out in this study that the use of combustion additions to eliminate visible emissions resulted in agglomerated particles of greater than 1 μ

TABLE 9-3

GAS TURBINE FUEL SPECIFICATIONS

Constituent	P&WA Spec. 527(1)	Suggested(2)	Westinghouse(3)	General Electric(1)
Sulfur	1.8 Mol % H_2S	<1 Mol % H_2S or Less Than Amount Required to Form 5 ppm Alkali Metal Sulfates	2% by Weight	Less Than Amount Required to form 3 ppm Alkali Metal Sulfates
Particulates	0.08 lb/10^6 ft^3 (0.00056 gr/ft^3)	0.01 gr/ft^3 40 μ Maximum	Limits by Material(4)	30 ppm (Weight) (0.01 g/ft^3)
Metals				
Vanadium	<0.2 ppm (Weight)	-----	0.5 ppm (Weight)	-----
Sodium and Potassium	<0.6 ppm (Weight)	see Sulfur Spec.	0.5 ppm (Weight)	See Sulfur Spec.
Calcium	0.1 ppm (Weight)	-----	10 ppm (Weight)	-----
Lead	0.1 ppm (Weight)	-----	2 ppm (Weight)	-----
Copper	0.2 ppm (Weight)	-----	-----	-----

(1) For aircraft-derived turbine using gaseous fuels

(2) For industrial turbines; subject to revision

(3) Liquid fuel specifications

(4) Given in Ref. 5-9 as 4×10^{-4} gr/ft^3 of 2μ to 10μ particulate in gaseous fuels

(above the size range which results in refraction). While deposition did occur, there was no erosion which indicates that the soft compounds which left the airstream and impinged on the blades did not damage them.

Tests in the United Kingdom on fluid-bed combustion (Reference 9-6) also indicate that soft particles with quite large diameters of about 50 μ do not appear to cause erosion. However, hard particles may result from coal combustion above the ash fusion point Tests carried out in Australia (Reference 9-7) indicate that such particles cause erosion at sizes above 6μ. Perhaps the most useful guide is work done in the petro-chemical industry on turbine erosion (Reference 9-8). Results of this work are shown in Figure 9-1, where engine life is shown as a function of particle size and loading. The shape of Figure 9-1 indicates that the relationship between lifetimes, particulate loading and particulate size would be of the form

$$L = \frac{\text{Constant}}{\mu \cdot C} \tag{9-1}$$

where

L = lifetime

μ = particle size

C = loading

Thus, for a given lifetime, a reduction in particle size would be accompanied by higher allowable loadings. For example, assuming linear particle size distribution, capture of particles of 2 μ and above would allow twice the particulate concentration compared to what would be allowed by capture of 4 μ and larger particles.

Suggested Low-Btu Fuel Gas Specifications

The values for loadings in Table 9-3 are based upon methane fuel. For low-Btu fuel, both the heating value and the density will affect the allowable loadings. For fuel gases with heating values of 150 Btu/ft^3 (about 2500 Btu/lb), the value in Table 9-3 would be reduced by approximately a factor of eight if the solids loading to the turbine were to be kept constant. Using the foregoing approach, Table 9-4 has been prepared to serve as a guideline for low-Btu fuel gas.

In addition to the usual fuel property limitations, Table 9-4 includes a value for fuel nitrogen compounds expresed in terms of NH_3. The limitation is based upon 90 % conversion of fuel nitrogen to NO_x.

SULFUR EMISSIONS

The variation in sulfur emission shown in the summary table is somewhat arbitrary. In most cases, changes in equipment design could be effected to

FIG. 9–1

EFFECT OF PARTICLE SIZE ON ENGINE LIFETIME

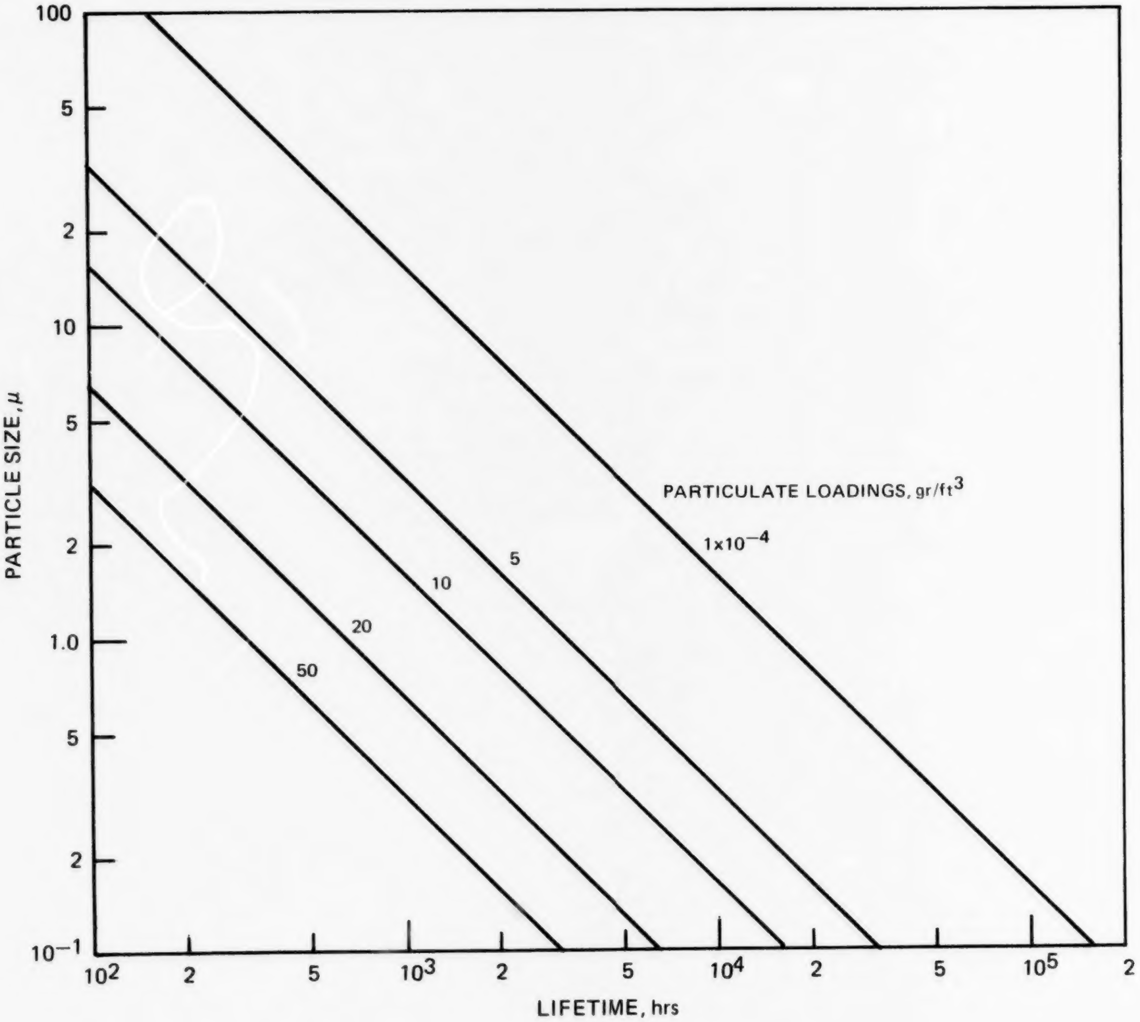

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TABLE 9-4

SUGGESTED LOW-BTU FUEL GAS CLEANUP SYSTEM GOALS

Property	Specification	Resulting Emission
Sulfur	0.05 Mol % or Less Than Amount to Form 0.6 ppm Alkali Metal Sulfates	0.44 lb SO_2/MMBtu
Particulates	4 ppm (weight) (0.0012 gr/ft^3)	<0.01 lb
Metals		
Vanadium	<0.03 ppm (weight)	
Sodium and Potassium	See Sulfur Spec	
Calcium	<0.012	
Lead	<0.012	
Copper	<0.0025	
Nitrogen Compounds	500 ppm as NH_3	0.3 lb NO_2/MMBtu

reduce the levels. Also, estimated raw gas compositions have a decided effect on cleanup performance. Such factors have been examined in an effort to identify the sensitivity of each design to sulfur emission level.

An important factor in all systems, especially as removal efficiency increases, is the form in which the sulfur appears in the raw fuel gas. In this study it has been assumed that all sulfur would be in the form of COS or H_2S; however, potentially significant quantities of other sulfur compounds could be present (as compared to a 100 ppm level in the clean gas) Since other compounds have not been definitively or quantitatively identified, their effect on the system has not been evaluated in any detail. While it is anticipated that the cleanup processes will be able to cope with such gases, removal or recovery performance will likely suffer. Thus, where sulfur emissions are shown to be on the order of 0.1 $lb/10^6$ Btu, it is quite possible that they could be higher by a factor of two or that the associated equipment may be more complex than anticipated.

Low-Temperature Systems

With the exception of the molten salt system, all the systems studied herein incorporate a Selexol desulfurizer with sulfur recovery via a Claus plant with Beavon tail-gas cleanup. Claus plant emissions are virtually zero and the emission level is ultimately determined by the Selexol system design. In the case of the molten salt gasifier, sulfur removal occurs both in the gasifier and in the CO_2 recovery tower which is a necessary part of the salt recovery process. The H_2S is stripped from the solution and sent to a Claus plant with tail gas cleanup as in the Selexol designs. Thus, in both cases, sulfur emissions depend only on the degree of removal of sulfur compounds from the main fuel gas stream.

Selexol Design Basis--

Each of the Selexol designs is based on removal of virtually all H_2S (down to approximately 30 ppm). Under these conditions, only 30% of the COS is removed. Thus, nearly all of the sulfur emissions are due to COS in the fuel gas. The level of COS in the raw gas would be on the order of 5% of total sulfur at equilibrium (see Tables 8-12 and 8-13). As hard data are generally unavailable, estimates vary between equilibrium and two times the equilibrium level. This is the principal reason for the variation in emission levels between the BCR and U-gas systems.

Variation of Sulfur Emission Level--

The parametric studies presented in Section 8 showed that, for a refrigerated Selexol absorber, there was little to be gained in terms of performance and cost by allowing the level of H_2S in the clean gas to rise from 30 to 1000 ppm. However, to further reduce total sulfur level in the gas will require a significant increase in solvent flow rate. This in turn will result in increased equipment cost and increased utility consumption.

When solvent flow rate is sized for H_2S removal, approximately one third of the COS is removed. To remove COS to the 100 ppm level requires that solvent flow rate be significantly increased. To achieve a high degree of removal of a particular component requires a solvent flow-rate inversely proportional to its solubility. The effect of this change on the performance and cost of two air-blown BCR-type gasifiers designs is compared in Table 9-5. Utilities increase by a factor greater than two, while cost is approximately doubled. Power output was penalized at the rate of 20 lbs steam/kWh giving a total decrease in plant output of some 38 MW in going from H_2S to COS based design. Of the $48/kW increase in fuel system cost, some $10/kW can be attributed to the decreased output.

The Selexol desulfurizer systems for the other gasifiers considered were sized on the basis of H_2S removal. Therefore, a similar improvement in COS removal can be expected for each of the other systems with a comparable increase in both cost and utility consumption. The conclusion to be drawn is that desulfurizer cost can vary by a factor of two depending on the composition of the raw fuel gas.and the degree of sulfur removal which is required.

Molten Salt System--

Because 85% of the sulfur is retained in the melt, total sulfur in the raw gas is quite low. The COS level is reduced correspondingly and, even with no COS removal, the emission level is quite low. After leaving the gasifier, the raw gas is scrubbed with a sodium carbonate solution to recover CO_2. During this process, most of remaining H_2S is removed. In this study, no COS removal is assumed (Reference 9-9). This may be a conservative estimate as COS is reported to be hydrolyzed (Reference 9-10) in the hot potassium carbonate solution of the Benfield system. Under appropriate conditions complete COS removal is claimed for that process.

NITROGEN OXIDES

Conventional gas turbine burners firing distillate fuels are concerned only with thermally generated NO_x. Its formation is both temperature and time dependent. Water injection is the current method used to control thermal NO_x emissions, but they could eventually be controlled by burner design. Fuel gas from coal can be expected to contain significant amounts of nitrogen bearing compounds (usually lumped together as ammonia). At the levels expected and under the expected burner design conditions, virtually all of the nitrogen in the ammonia will form nitric oxide (Ref. 9-11). It so happens that measures most appropriate for reduction of thermal NO_x tend to enhance the production of NO from ammonia. Because of this, both problems will be treated separately. Low-temperature systems which include a water wash need be concerned only with thermal NO_x. In the high-temperature systems, a choice must be made between further fuel processing and combustion modification.

TABLE 9-5

COMPARISON OF SELEXOL DESIGNS
BCR Air-Blown Gasifier

	H_2S Based Design	COS Based Design
L. P Steam Required - lb/hr	131,500	317,000
Power Required - kW	20,400	48,800
Installed Cost - $\$10^6$	26.6	51.1
H_2S in Clean Gas - ppm	32	10
COS in Clean Gas - ppm	450	90
Sulfur Emission - lb SO_2/MMBtu	0.39	0.08
Net Plant Output - MW	1,098	1,060
Fuel System Cost - $\$10^6$ (Including Contingency, Escalation and Interest During Construction)	277.5	319.1
Specific Cost - $/kW	253	301
Generating Efficiency -	0.439	0.424

Thermally Produced Nitrogen Oxides

Data for various types of fuel gas in conventional burners show that low-Btu fuel gas will produce little thermal NO_x. The range of data for low-, medium-, and high-Btu gas with fuel at ambient temperatures is given in Figure 9-2. However, reheating of the clean fuel gas, while improving efficiency, results in a higher flame temperature and, thus, more thermal NO_x. As Figure 9-2 shows, burning at off-stoichiometric conditions, i.e., at lower than stoichiometric flame temperatures, offers a means of controlling production of thermal NO_x. The data in this figure were obtained by Turbo Power and Marine Systems a subsidiary of United Technologies Corporation. They have been carrying out a series of tests on low- and medium-Btu fuel gases produced by an experimental gasifier at the Texaco Development Company's Montebello, California research facility.

Of particular interest are results from testing of premixed (fuel and air mixed prior to introduction into the combustor) burners. Figure 9-2 (Reference 9-12) shows the NO_x emissions as a function of source temperature rise for several values of fuel chemical heating value. In the test series, the fuel gas was delivered at essentially ambient temperature. The theoretical stoichiometric temperatures are given. In conventional burners, indicated by the areas between the solid lines, stoichiometric temperatures are generally achieved. Also shown in Figure 9-2 are the approximate emissions for burners in which the fuel and air have been premixed. These are shown as off-stoichiometric conditions since complete mixing will eliminate the possibility of stoichiometric combustion which occurs in conventional burners. While no significant reduction was noted for low-Btu gas (the emissions were already low), use with the medium-Btu gas indicated a large reduction.

Stoichiometric flame temperatures given in Figure 9-2 apply to the point of maximum burner ΔT. Reduced burner ΔT corresponds to part throttle conditions for a gas turbine. At part throttle conditions, combustion air temperature is decreased with a corresponding decrease in stoichiometric temperature and NO_x production. As flame temperature is increased above 3600 R, the increase in thermally produced NO_x is exponential. For the systems studied here, fuel gas temperatures ranged from 800 to 1000 F and the resultant stoichiometric flame temperatures, given in Table 9-6, are in the high thermal NO_x zone. Figure 9-3 (Reference 9-13) gives an estimate of the concentration as a function of combustor exit temperature and theoretical flame temperature. It differs from the data of Figure 9-2 in that it shows a stronger temperature dependence. However, it is clear from both figures that conventional combustors will result in the production of excessive thermal NO_x. (Note that the data of Figure 9-2 is at atmospheric pressure and must be corrected for turbine pressure ratio.).

FIG. 9–2

NO_x PRODUCTION FROM COMBUSTORS BURNING LOW–BTU AND MEDIUM–BTU GAS

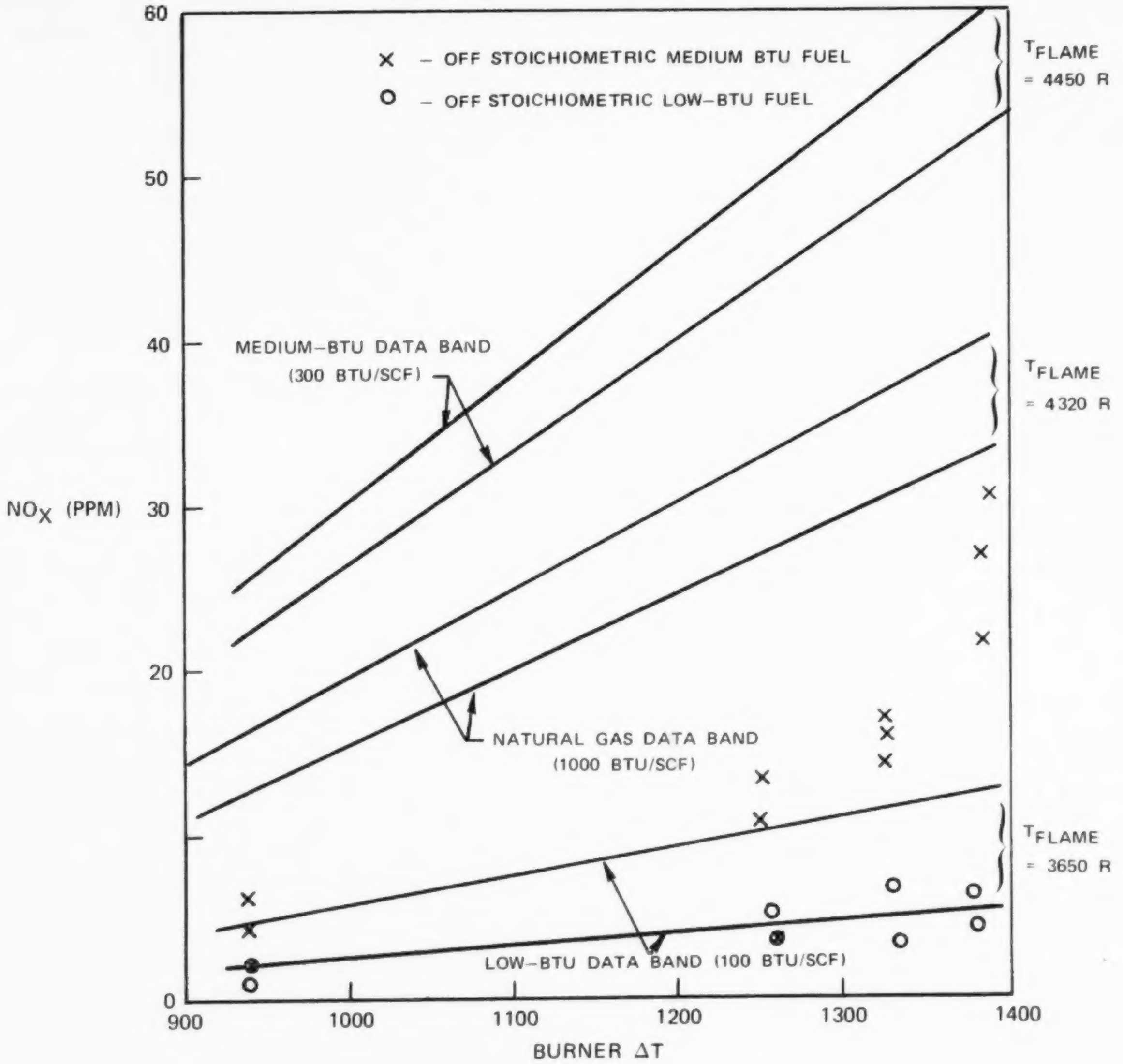

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TABLE 9-6

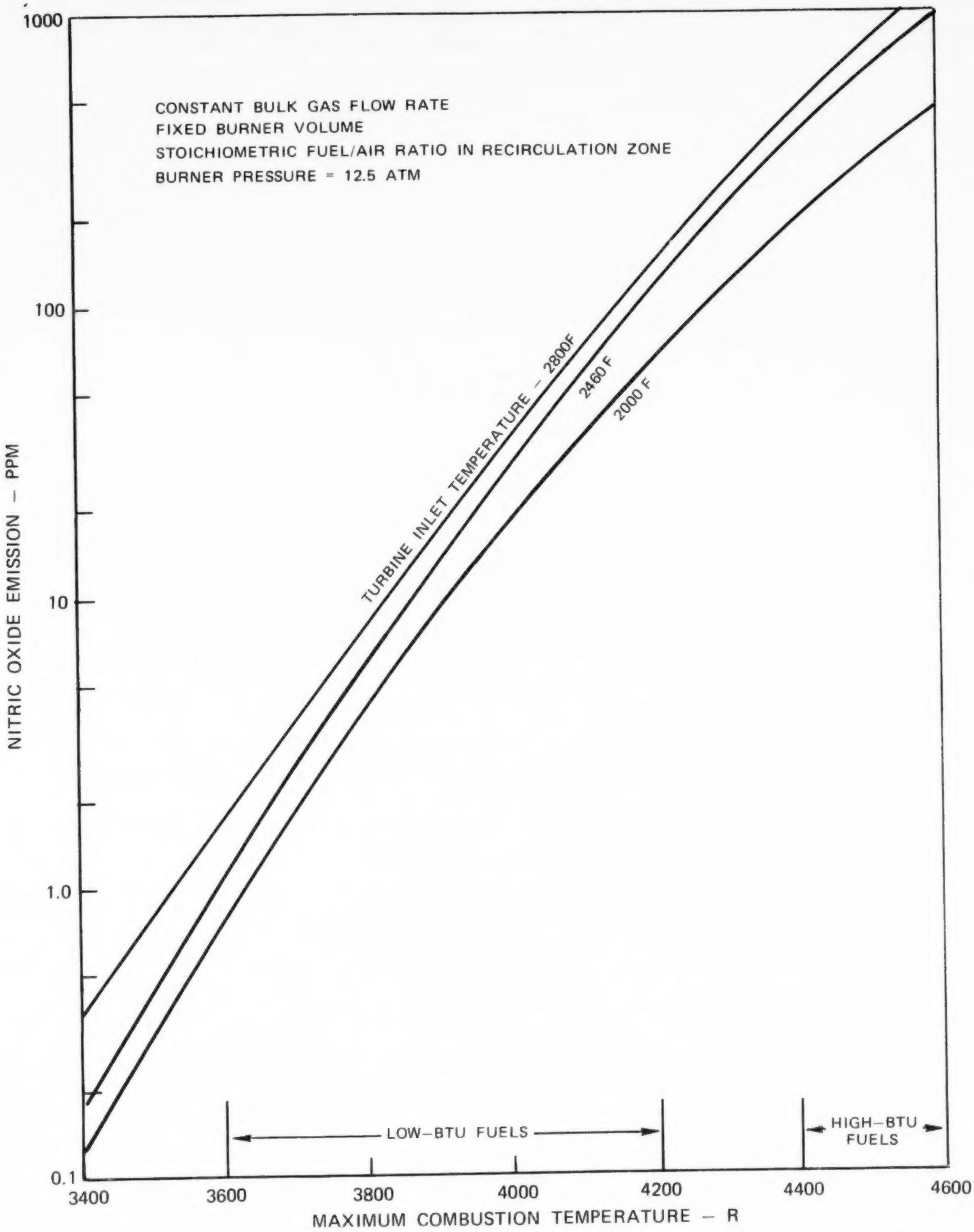
THEORETICAL FLAME TEMPERATURES

Air at 793 F
Equivalence Ratio = 1.0

System	Fuel Gas HHV Btu/SCF	Fuel Gas Temp F	Fuel/Air lb/lb	Flame Temperature F/R
U-Gas Selexol	158.6	944	.681	3734/4194
U-Gas Low-Steam Selexol	169.7	944	.639	3870/4330
BCR Air-Blown Selexol	177.9	926	.608	3946/4406
BCR O_2-Blown Selexol	332.4	926	.262	4494/4954
Molten Salt	154.0	805	.724	3800/4260
U-Gas Low-Steam Conoco	160.9	1,000	.675	3805/4265
BCR Air-Blown Conoco	170.0	1,000	.639	3900/4360
BCR O_2-Blown Conoco	268.0	1,000	.335	4240/4700

FIG. 9–3

NITRIC OXIDE FORMATION IN GAS TURBINE BURNER



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Thermal Formation Mechanism--

Complex computer simulations have been developed by United Technologies and others that model the combustor internal flow-field, the combustion reactions, and the thermal NO_x kinetics (Reference 9-14 through 9-19). The simplified NO_x kinetic predictive techniques are generally based on the fact that the NO formation rate is very slow relative to the hydrocarbon combustion reaction rate so that the two can be decoupled in predicting NO formation; i.e., the combustion reactions can be assumed to be at equilibrium in estimating NO formation rates. This is illustrated in Fig. 9-4 which is the result of a kinetic model considering both combustion and NO kinetics (Reference 9-20). For that system, the combustion reaction is essentially complete in 60 microseconds at which time the NO concentration is several orders of magnitude less than its final or equilibrium value.

The thermal mechanism for formation of NO from nitrogen and oxygen was originally proposed by Zeldovich (Reference 9-21). It consists of a chain of two reactions for the production of nitric oxide:

$$O + N_2 \rightleftarrows NO + N \tag{9-2}$$

$$N + O_2 \rightleftarrows NO + O \tag{9-3}$$

While there is evidence of the formation of "prompt" NO (Reference 9-22), it is believed that the bulk of the NO is formed by the thermal mechanism. A third reaction,

$$OH + N \rightleftarrows NO + H \tag{9-4}$$

can become important when the oxygen concentration becomes low under fuel-rich conditions. Several equivalent solutions to the above equations are given in References 9-23 through 9-25. They all assume that the nitrogen atoms are in equilibrium with the products of combustion and that the flame temperature is constant.

By making several approximations, a simplified solution (Reference 9-25) can be obtained that shows good agreement with more complex models and correlates with test data:

$$\rho = \frac{\theta t}{2} \tag{9-5}$$

FIG. 9–4

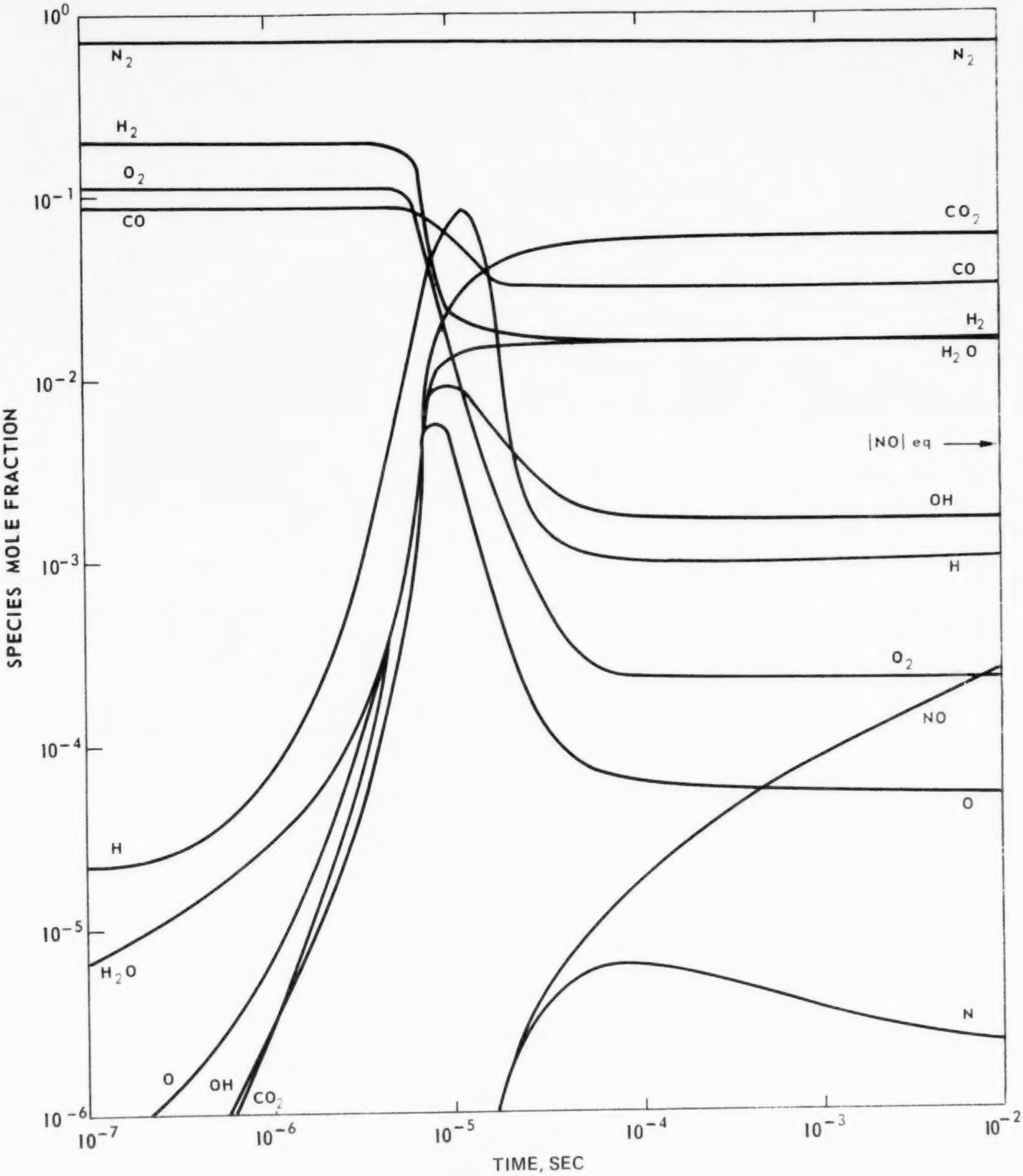
CONCENTRATION–TIME PROFILES FOR PREMIXED H_2–CO–AIR MIXTURE

INLET TEMPERATURE = 1800 R

PRESSURE = 10 ATM

COMBUSTION TEMPERATURE = 4460 R

EQUIVALENCE RATIO = 1.0



where $\rho = \dfrac{X_{NO}}{(X_{NO})_e} = \dfrac{\text{mole fraction NO}}{\text{mole fraction NO at equilibrium}}$

t = apparent residence time, sec

and $$\theta = \frac{4.24 \times 10^{15} P^{1/2}}{T} (X_{N_2})^{1/2} \exp \frac{-114{,}572}{RT} \qquad (9\text{-}6)$$

where θ = NO formation parameter, sec^{-1}

T = temperature, K

P = pressure, atm

X_{N_2} = mole fraction of N_2

R = gas constant, 1.987 $\dfrac{cal}{mole - {}^oK}$

As suggested in Reference 9-23, if X_{N_2} is assumed to be constant, the curve shown in Figure 9-5 would result. The mole fraction of N_2 is in the range of .71 to .73 for distillate and low-Btu gas. It can be as low as 0.6 for medium-Btu gas but this only introduces a 10% error. The above are a close approximation for values of ρ on the order of 0 to 0.2 and equivalence ratios near unity. Equivalence ratio is defined as the ratio of actual fuel/air ratio and the stoichiometric fuel/air ratio.

Equilibrium composition for a low-Btu fuel is shown in Figure 9-6. In estimating NO_x formation, the parameters of interest in Equation 9-5 are equilibrium NO concentration and adiabatic flame temperature. These are presented in Figures 9-7 and 9-8 as a function of equivalence ratio and fuel temperature. While fuel temperature does have a significant effect on flame temperature and rate constant, it is also apparent that adjustment of the equivalence ratio (off-stoichiometric combustion) could be valuable in offsetting their bad effects.

Reduction of Thermal NO_x--

The NO_x reduction potential for off-stoichiometric combustion is shown in Figure 9-9 for two modern aircraft engines burning liquid fuel. Data for methane with air in a well stirred reactor are also shown. All data have been corrected to similar conditions to permit comparison. It appears that the practical limit that can be achieved as well as the ability of the combustor to achieve that limit tend to favor lean conditions ($\phi < 1.0$).

FIG. 9–5

NO RATE PARAMETER AS A FUNCTION OF TEMPERATURE

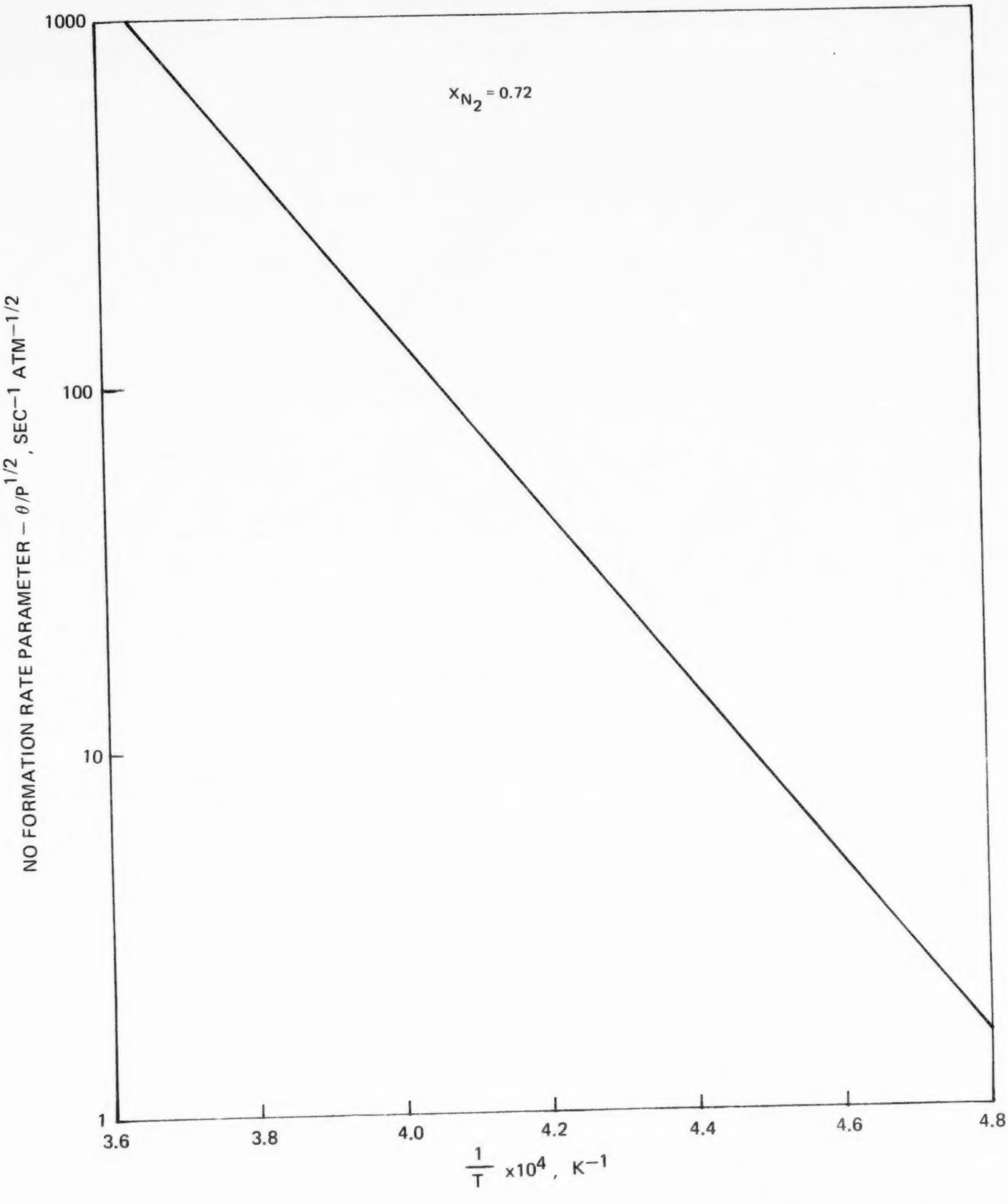


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FIG. 9–6

EQUILIBRIUM COMBUSTION PRODUCTS LOW BTU GAS
144 BTU / SCF

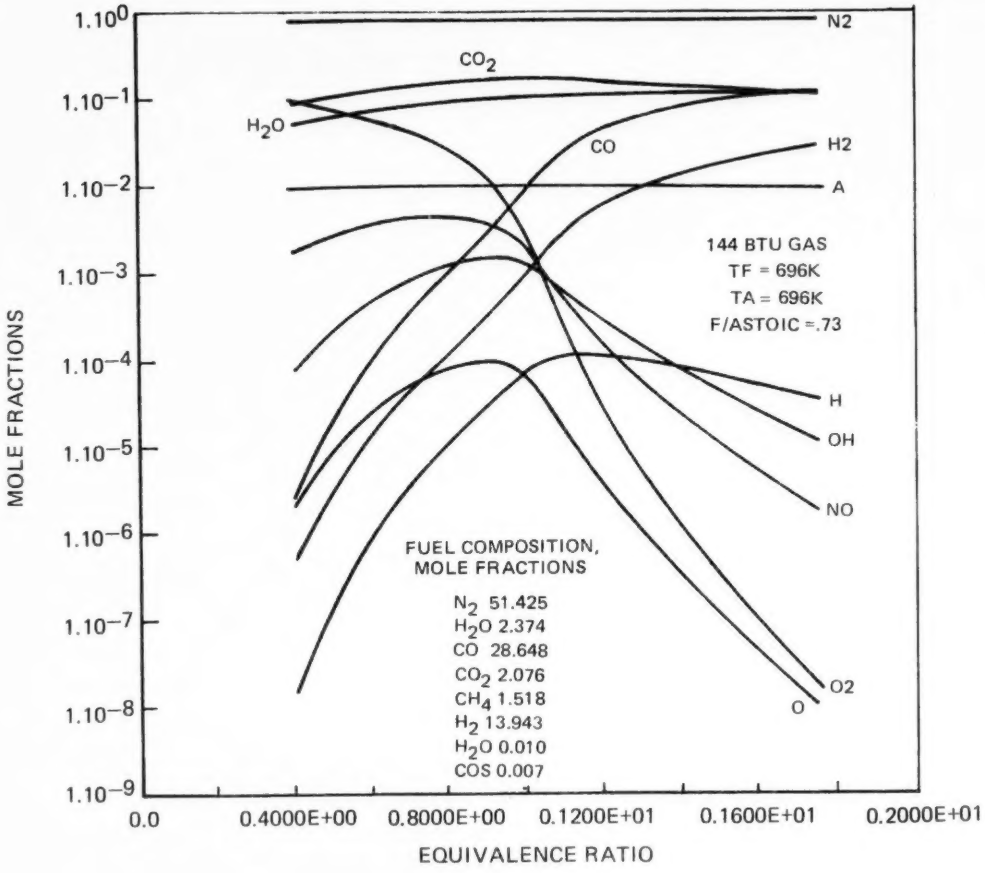
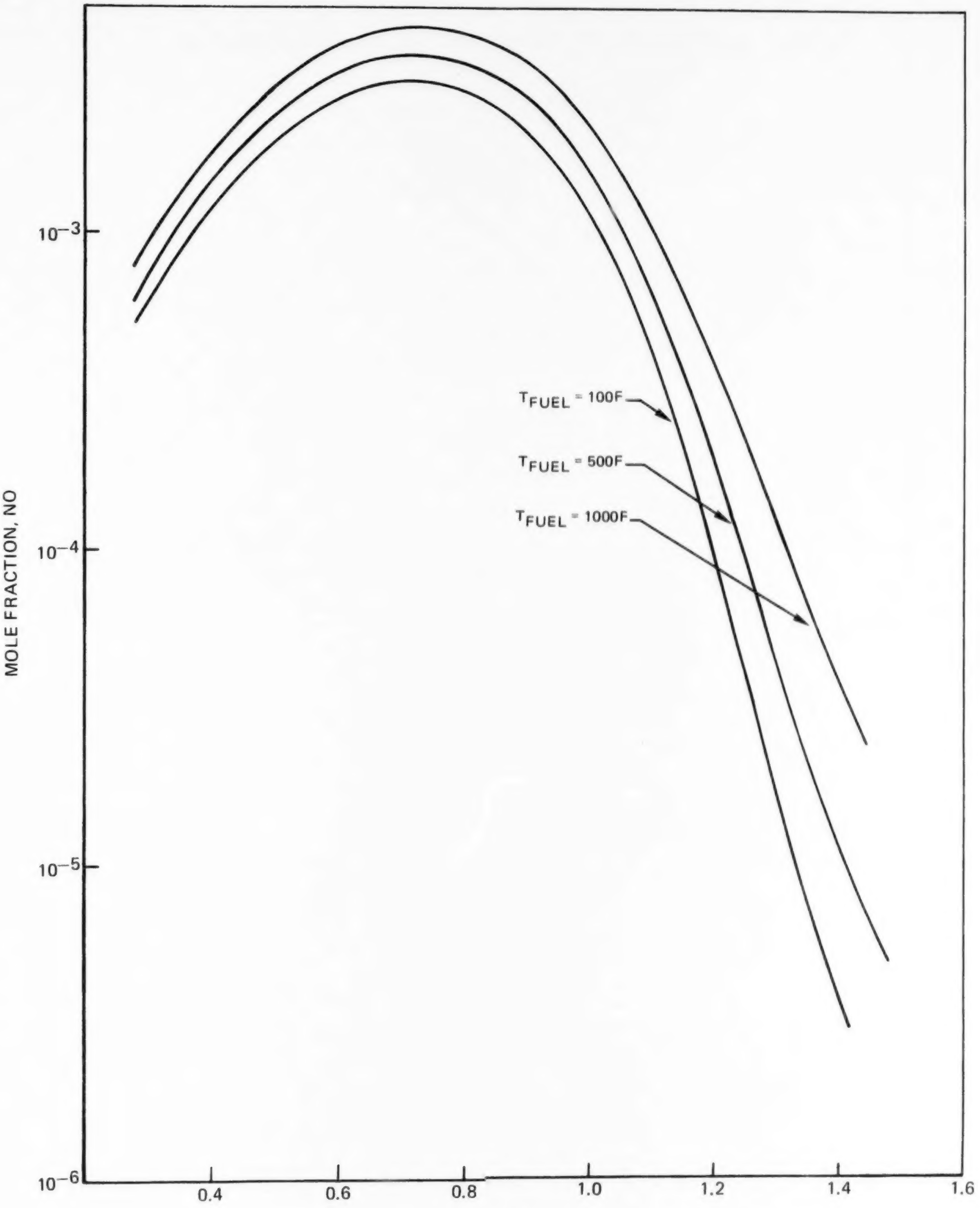


FIG. 9–7

EQUILIBRIUM NO CONCENTRATION MOLTEN SALT GASIFIER FUEL



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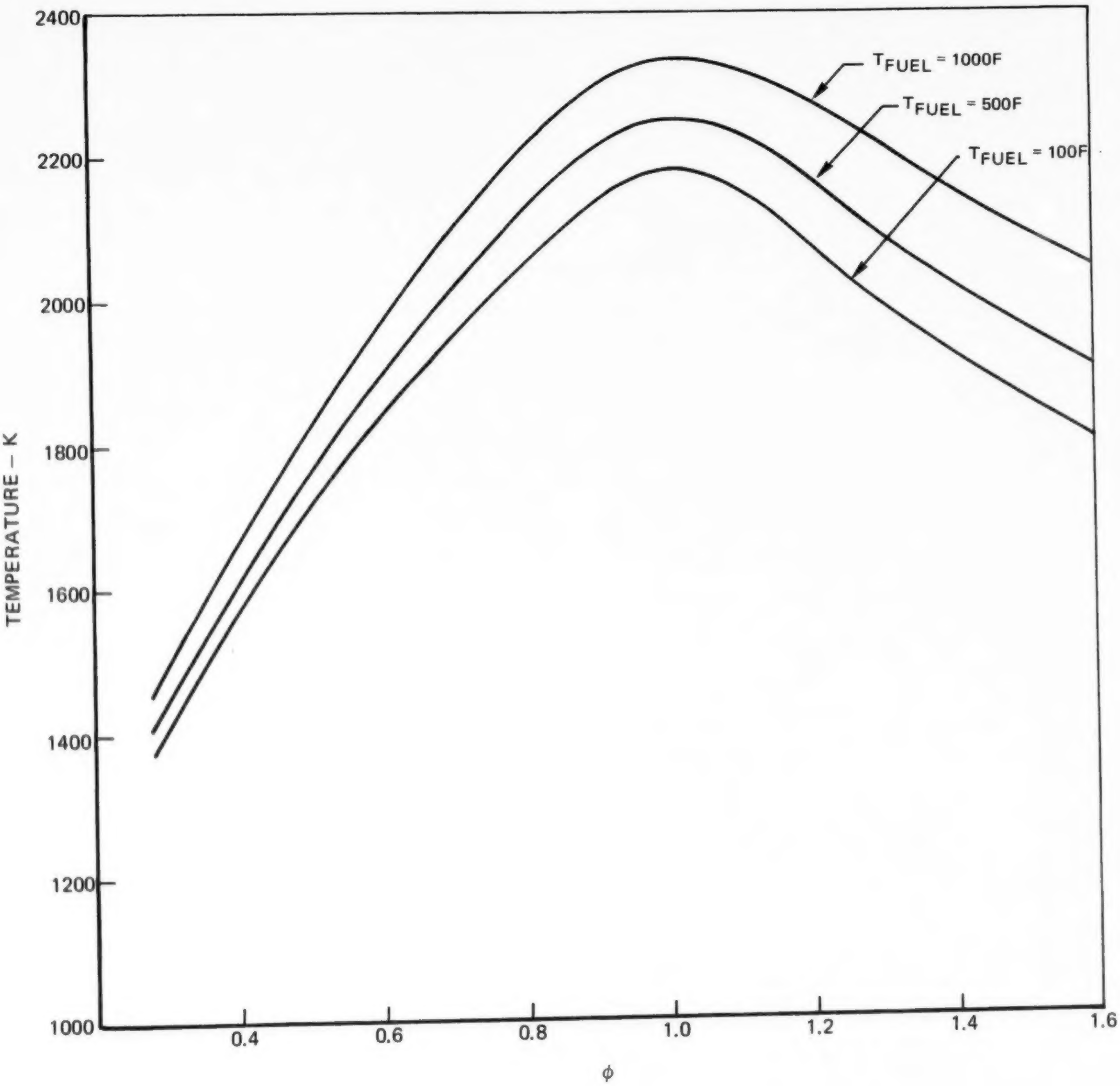
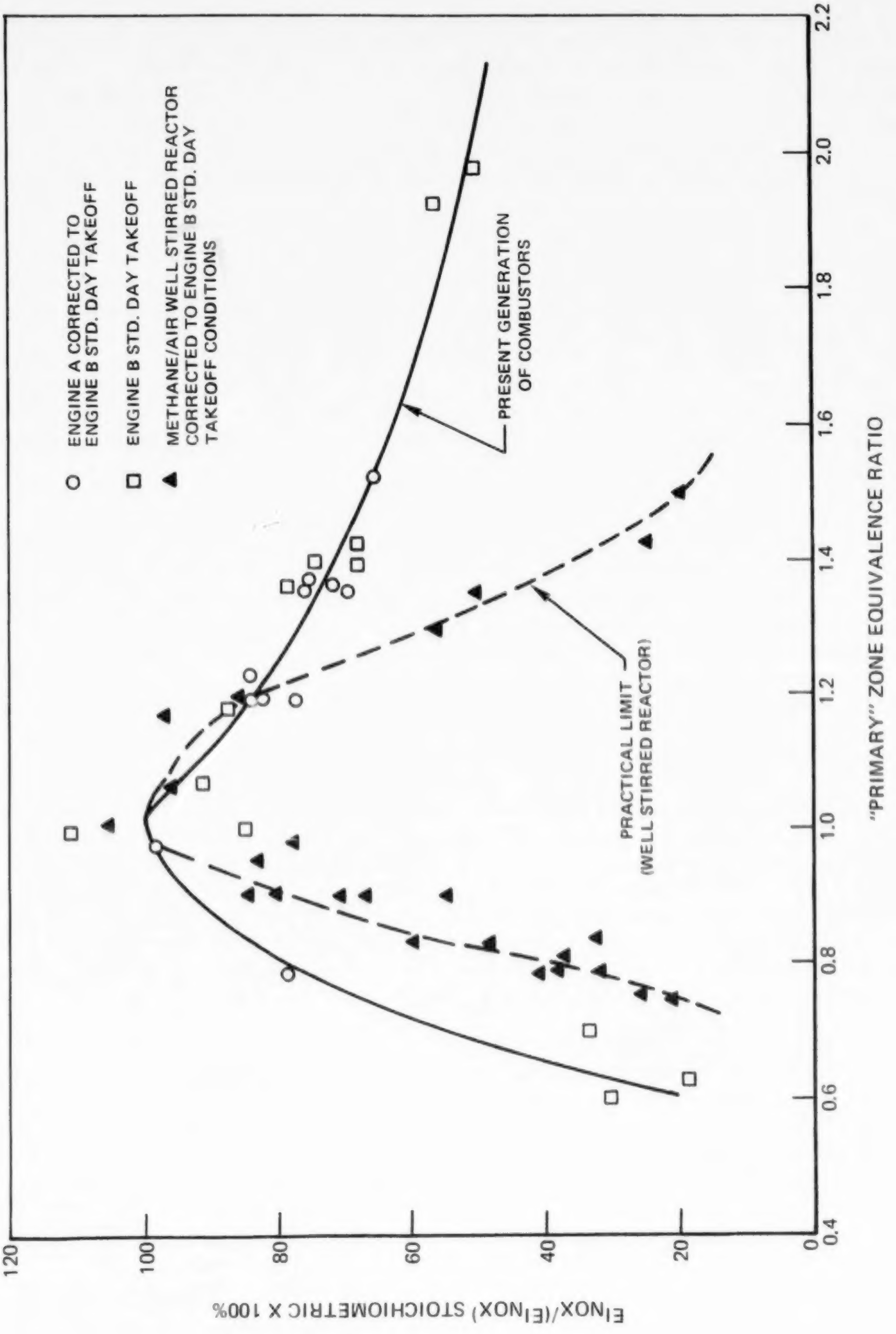
ADIABATIC FLAME TEMPERATURE MOLTEN SALT GASIFIER FUEL



78–02–157–4

FIG. 9–9

COMPARISON OF CONVENTIONAL COMBUSTOR NO_x–EQUIVALENCE RATIO RELATIONSHIP WITH THAT OF A WELL–STIRRED REACTOR



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Comparing the engine results with those of the well-stirred reactor reveal the very important effects on NO_x formation of fuel preparation and mixing. With gaseous fuel, the well-stirred reactor probably represents the best level of mixing that can be achieved in practice without premixing the fuel and air prior to combustion. For the benefits of mixing to be worhwhile, the conclusion is inescapable: if either a rich or a lean combustion alone is selected to reduce NO_x, considerably better mixing will be required than is presently obtainable with conventional liquid-fueled combustors.

The reduction in NO_x that can be expected from reducing residence time in the flame zone for lean combustion systems is shown in Figure 9-10. The data shown represents relatively unsophisticated "early quench" methods where large amounts of relatively cool dilution air are introduced very early, at or before the end of the primary zone. This is done to freeze chemical reactions. As can be seen, to achieve significant reductions in NO_x with this method alone, very large reductions in residence time are required. In general, the required reductions cannot be achieved without disturbing combustor performance.

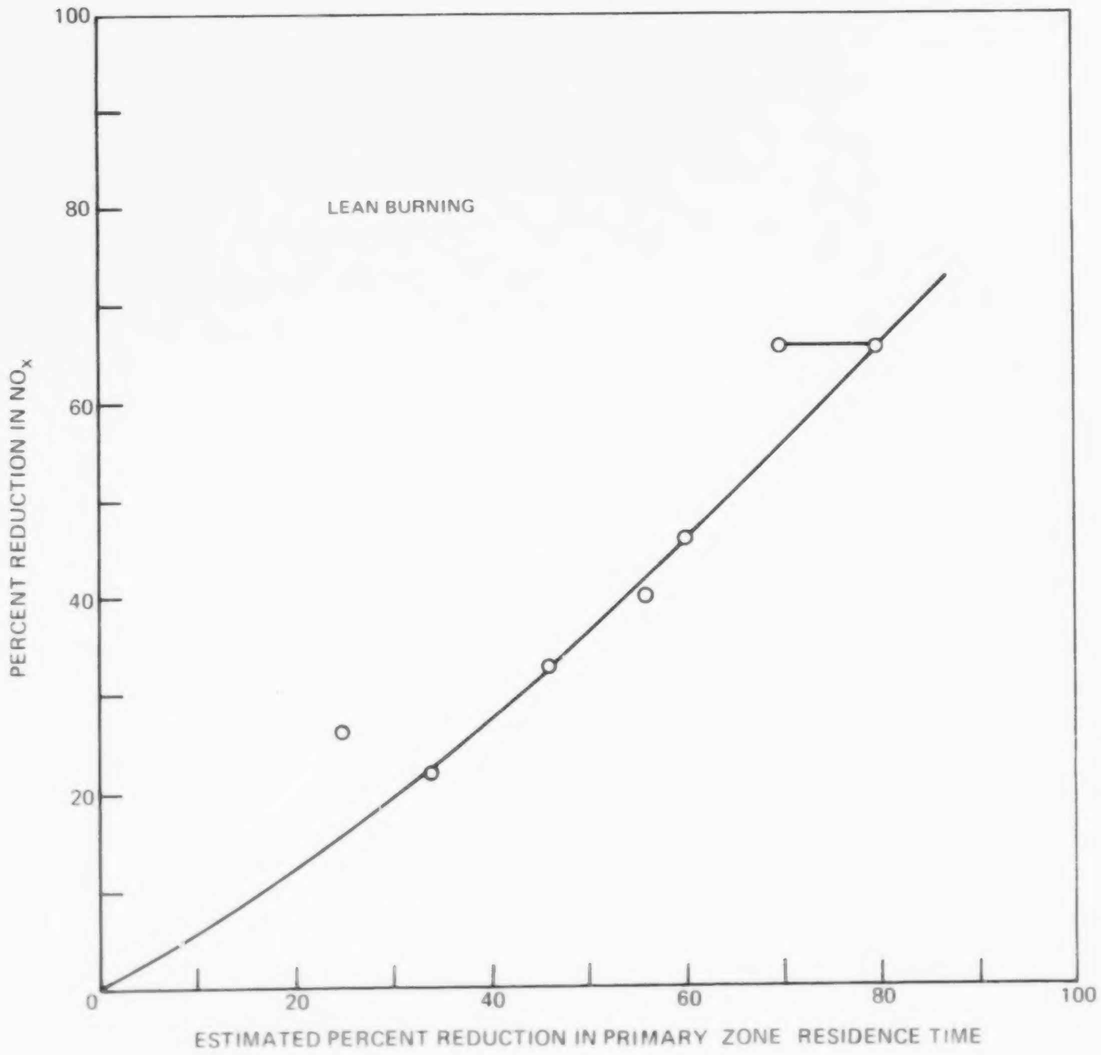
Mixing on a molecular level is necessary for the combustion of a mixture to take place at a flame temperature equivalent to the bulk fuel-air ratio. Typical characteristic times, shown in Table 9-7 for combustion of natural gas in air in the FT4 combustor, indicate that mixing time is the largest characteristic time of those for the processes involved. Therefore, if the full benefit of low flame temperatures were to be realized by burning off-stoichiometric, premixing of the reactants prior to introduction into the combustor would be necessary. If this were not done, combustion would be controlled by mixing and diffusion flames would exist with their associated high temperatures.

When good premixing is achieved, large reductions in exhaust NO_x can be obtained. Figure 9-11 shows the arrangement of a combustor rig in which premixing to an equivalence ratio of 0.75 was achieved external to the combustor in a high-pressure drop mixing chamber. The fuel was natural gas and the test pressure was 1.3 atmospheres. The residence time after mixing was limited to 8 msec to avoid autoignition. The fuel and air mixture was introduced into a modified FT4 combustor through a multitude of individual tubes. Residence time in the primary zone was estimated at 3 msec.

The premixed results are compared in Figure 9-12 with the results from a conventional FT4 combustor when burning No. 2 liquid fuel at the conditions for the rig tests. The results of an early quench to about the same bulk equivalence ratio are also shown. The low NO_x benefits of premixing the fuel and air prior to combustion are clearly revealed. The limited mixing available in the conventional combustor is demonstrated by the early quench results which, although showing a significant reduction in NO_x, do not show anything close to the potential of the premix rig.

FIG. 9–10

REDUCTION IN NO_x EMISSION BY REDUCTION IN RESIDENCE TIME–INFORMATION BASED ON EXPERIMENTAL DATA



78–02–105–3

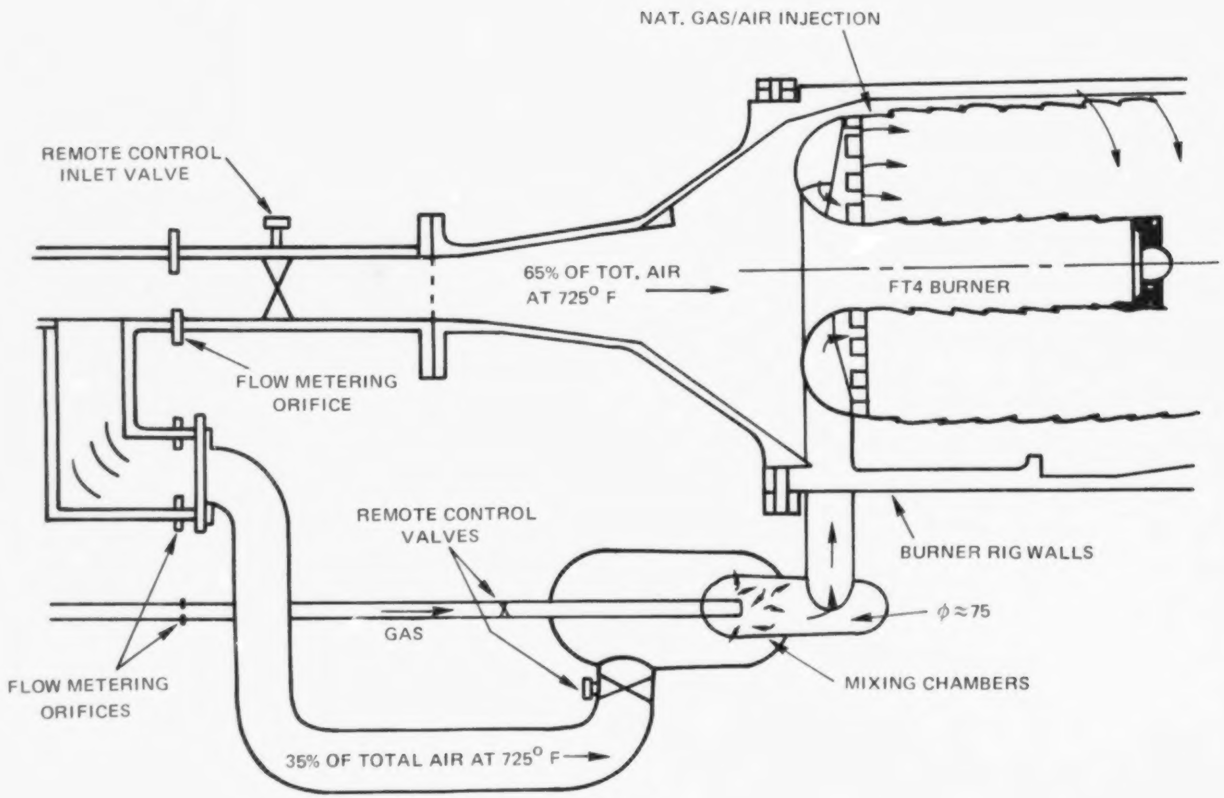
TABLE 9-7

CHARACTERISTIC TIMES FOR A NATURAL GAS COMBUSTOR

PRIMARY EQUIVALENCE RATIO	VAPORIZATION TIME, ms	TURBULENT MIXING TIME, ms	COMBUSTION TIME, ms	NO_x FORMATION TIME TO 500 ppmV, ms	RESIDENCE TIME, ms
0.7	0	10	0.20	19	5
1.0	0	10	0.13	0.7	5
1.2	0	10	0.2	106	5

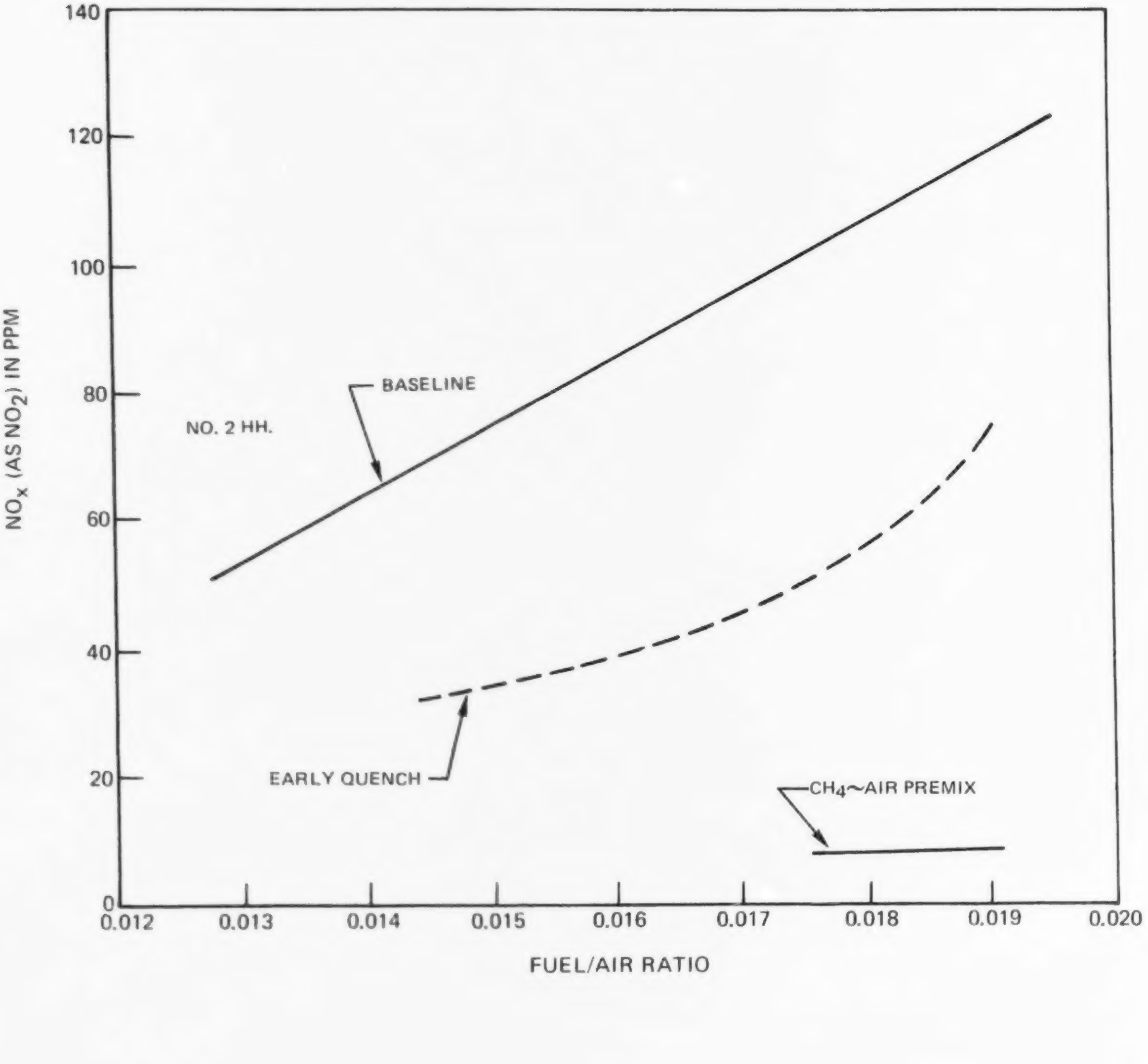
FIG. 9–11

FT4 PREMIX RIG WITH EXTERNAL MIXING


NAT. GAS/AIR INJECTION
REMOTE CONTROL INLET VALVE
65% OF TOT. AIR AT 725° F
FT4 BURNER
FLOW METERING ORIFICE
REMOTE CONTROL VALVES
BURNER RIG WALLS
GAS
$\phi \approx 75$
FLOW METERING ORIFICES
MIXING CHAMBERS
35% OF TOTAL AIR AT 725° F

78–02–105–4

COMPARISON OF RAPID QUENCH AND PREMIX NO_x WITH BASELINE



The effects of both residence time and equivalence ratio are shown in Figure 9-13 based on the Zeldovich mechanism. As stated earlier, it is a close approximation only near $\phi = 1.0$ (0.8 to 1.2). However, for preliminary design purposes it is considered to be adequate. While it would be feasible to design a conventional diffusion flame burner, studies at United Technologies (Reference 9-25) have shown that NO_x control would severely constrain design parameters and compromise other desirable combustion performance characteristics.

Premix Concept Design Considerations--

If the gaseous fuel can be thoroughly mixed with the combustion air on a molecular level prior to combustion, the flame temperature will correspond to that appropriate to the mixture bulk equivalence ratio. In addition, since the mixing is achieved external to the combustor, mixture residence time in the flame zone can be reduced. These two factors enable such a premixed system to take full advantage of the emissions reduction capability shown in Figure 9-13. This is done by operating the primary zone lean with a mean equivalence ratio about 0.8, and reducing the residence time in the primary zone to a minimum.

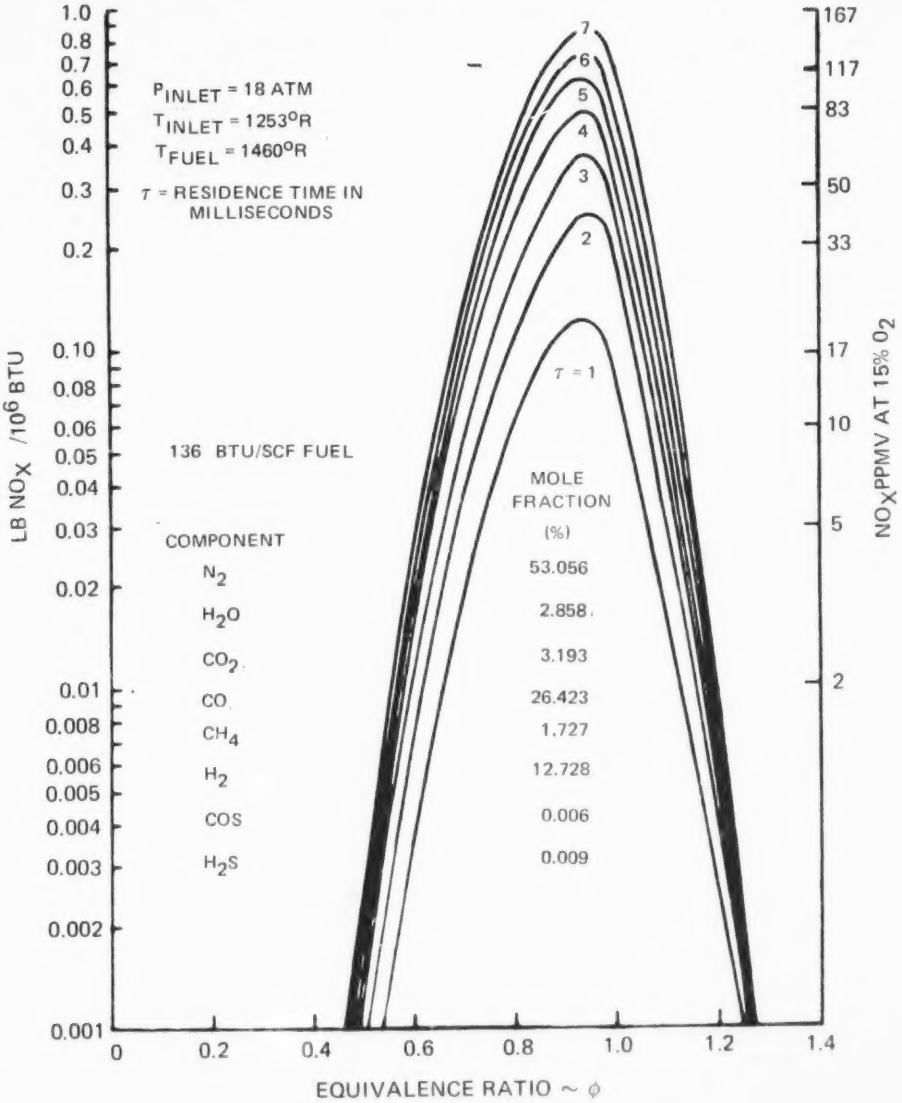
A preliminary design for a low-Btu gas combustor using this concept was prepared as part of the Department of Energy-sponsored High Temperature Turbine Technology Program (HTTTP) conducted by the Power Systems Division of United Technologies and UTRC (Reference 9-26). this preliminary design was based on the results of the joint P&WA-Texaco-Montebello test program for the combustion of gasified residual fuel oil in an FT4 combustor.

The concept is illustrated in Figure 9-14. Gaseous fuel is supplied to a manifold formed in the head of the combustor. Passing through and sealed in the manifold is an array of premix tubes about one inch in diameter which are open at their forward end to admit air entering from the diffuser section. Fuel from the manifold is admitted to individual tubes in this array through a number of metering ports arranged at the forward end of each tube in order to maximize the mixing length available. In the premix tube, the entering fuel jets mix with the air flowing through the tubes. This mixing is enhanced by turbulence grids attached to the premix tube inlets. The pressure drop across these grids also serves to inhibit fuel from flowing forwards out of the tube inlets into the air-casings in the event of a disturbance in the smooth supply of air to the premix tubes.

The premixed fuel and air are discharged from the premix tubes into the primary zone where combustion takes place. The downstream portion of the dome is formed into an air manifold for cooling purposes. Cooling air is admitted into the air manifold through ports on its sides. The premix tubes pass completely through the cooling air manifold. Cooling air from the manifold is discharged as a concentric annular jet from small annular gaps in the faceplate around the end of each premix tube. This air curtain so formed also serves as a flashback inhibitor, as will be described below.

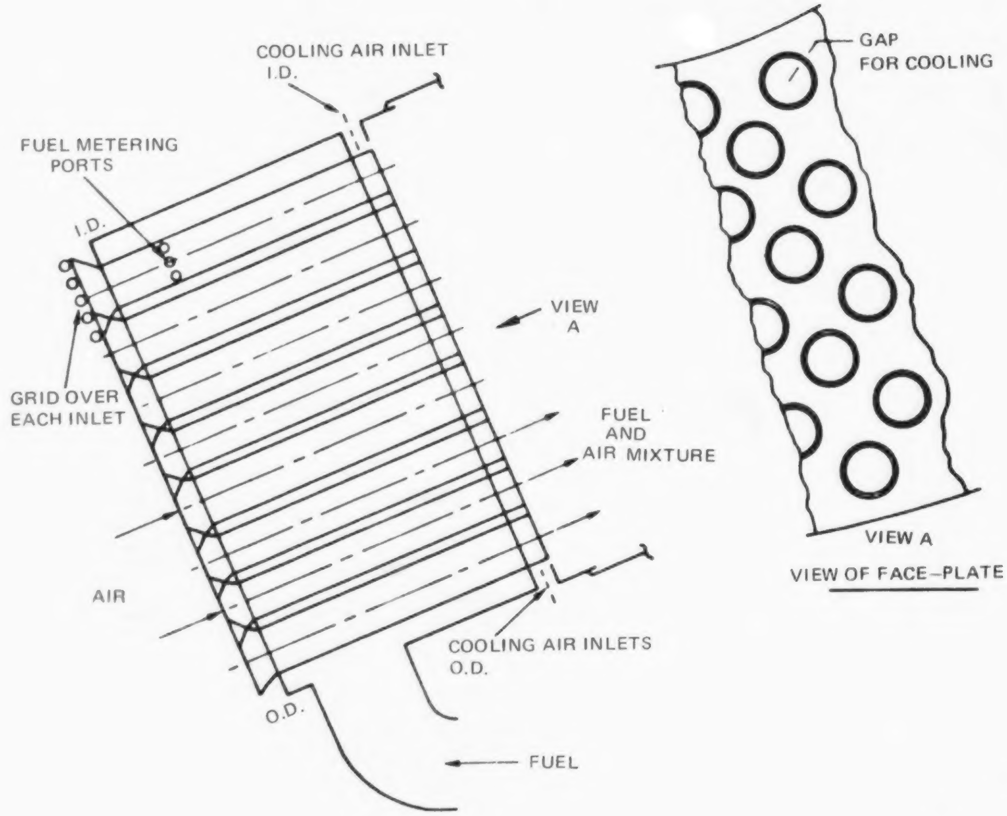
FIG. 9–13

NO$_X$ ESTIMATION FOR PREMIXED COMBUSTION BASED ON EXTENDED ZELDOVICH MECHANISM



COMPONENT	MOLE FRACTION (%)
N_2	53.056
H_2O	2.858
CO_2	3.193
CO	26.423
CH_4	1.727
H_2	12.728
COS	0.006
H_2S	0.009

SCHEMATIC OF PREMIX CONCEPT APPLIED TO A COMBUSTOR DOME


COOLING AIR INLET
I.D.
FUEL METERING
PORTS
I.D.
GRID OVER
EACH INLET
AIR
O.D.
VIEW
A
FUEL
AND
AIR MIXTURE
COOLING AIR INLETS
O.D.
FUEL
GAP
FOR COOLING

VIEW A

VIEW OF FACE–PLATE

FIG. 9–14

Further cooling air for the faceplate itself is discharged into the primary zone from the manifold through many "transpiration" cooling holes in the faceplate.

Figure 9-15 shows a diagrammatic representation of the flame stabilization in the low velocity regions of the jets of the premix concept. Again, this is a simplified diagram and does not attempt to show the additional flameholding which takes place in the recirculation zones on the faceplate between individual jets. The path length for fuel consumption is indicated. It will depend upon mixing tube diameter and rate of spread of the jet.

Autoignition--All premixed combustion devices are susceptible to a degree of risk from the dual hazards of autoignition and flashback. Autoignition, as strictly defined, is the temperature at which a fuel and oxidant mixture will spontaneously ignite when kept for an infinite residence time. When applied to flowing systems, it occurs when the residence time exceeds the ignition delay of the mixture. Flashback occurs when the flow velocity of the arriving combustible fuel and air mixture at some point in a flowing system becomes smaller than the burning velocity, and an external combustion wave is then able to propagate against the mixture stream into the supply duct. Both hazards, either occurring separately or together, can result in destructive burning in the supply duct, which might lead to secondary failures in the combustor and downstream components.

Using the collected data shown in Figure 9-16, the possibility of autoignition in the HTTTP-type design was evaluated. Provided there is not significant heat transfer to the mixture in the premix tubes, it was concluded that autoignition should not be a problem. Such heat transfer is not likely with careful design. However, Figure 9-16 shows the definite limitations imposed on heat transfer and fuel temperature by autoignition.

Flashback--Flashback is usually encountered in boundary layer and wake regions of a flow where local low velocities exist. It is usual to relate flashback to the boundary layer through a critical boundary velocity gradient. For turbulent pipe flow, the velocity gradient is given (Reference 9-27) by:

$$\frac{dU}{dy} = \frac{0.023\ \rho^{0.8}\ U^{1.8}}{(2R)^{0.2}\ \mu^{0.8}} \tag{9-7}$$

where:

R = tube radius

U = mean velocity of mixture flow through tube

DIAGRAMMATIC REPRESENTATION OF PREMIX COMBUSTOR

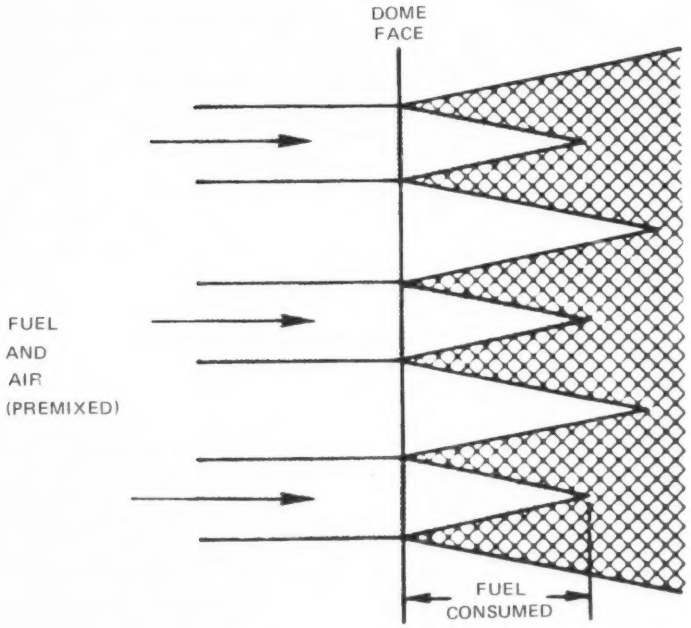
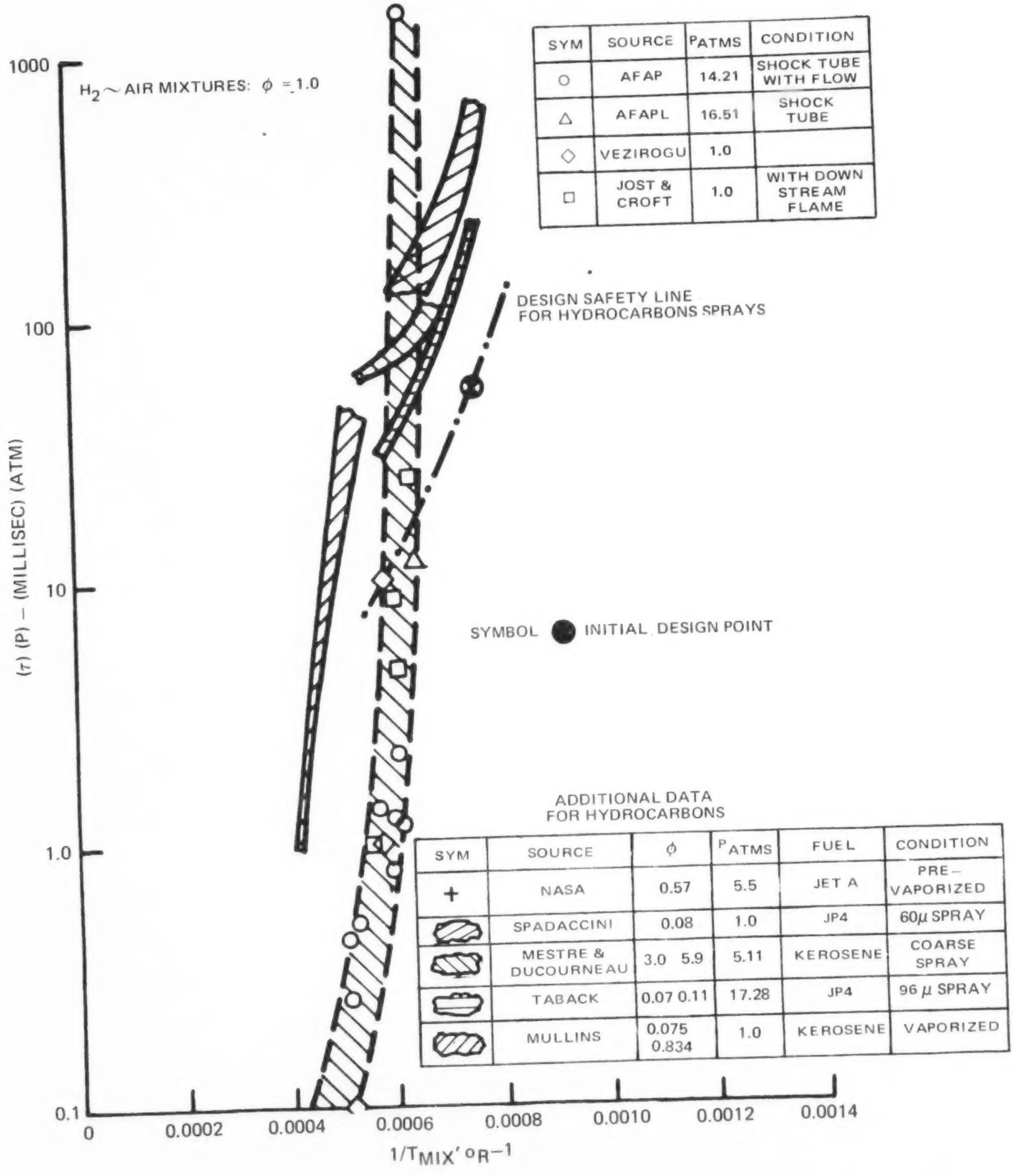

DOME
FACE
FUEL
AND
AIR
(PREMIXED)
FUEL
CONSUMED

FIG. 9–16

AUTOIGNITION CHARACTERISTICS–RESIDENCE TIME IN THE PRESSURE TUBE DUE TO LOCAL FLOW RECIRCULATIONS IS NOT LIKELY TO CAUSE A PROBLEM



SYM	SOURCE	P_{ATMS}	CONDITION
○	AFAP	14.21	SHOCK TUBE WITH FLOW
△	AFAPL	16.51	SHOCK TUBE
◇	VEZIROGU	1.0	
□	JOST & CROFT	1.0	WITH DOWN STREAM FLAME

ADDITIONAL DATA FOR HYDROCARBONS

SYM	SOURCE	ϕ	P_{ATMS}	FUEL	CONDITION
+	NASA	0.57	5.5	JET A	PRE–VAPORIZED
▨	SPADACCINI	0.08	1.0	JP4	60μ SPRAY
▨	MESTRE & DUCOURNEAU	3.0 5.9	5.11	KEROSENE	COARSE SPRAY
▨	TABACK	0.07 0.11	17.28	JP4	96 μ SPRAY
▨	MULLINS	0.075 0.834	1.0	KEROSENE	VAPORIZED

The value of dU/dy, at which flashback takes place is termed g_{crit} and is plotted against mixture bulk equivalence ratio to form a closed loop within which flashback takes place. Since burning velocity is a function of temperature and pressure, a series of such loops exists for a given fuel according to the test conditions.

Figure 9-17 presents estimated flashback loops for the preliminary design burner and for the premix Montebello testing (Reference 9-28). Also shown is the operating range of critical velocity gradient for the Montebello tests. For convenience they are shown at a single bulk value of hydrogen in the fuel, although in reality the bulk value varied. The operating range is to the left of and outside the flashback loops.

The Montebello testing could have been subject to incipient flashback since the operating range overlaps the estimated flashback loops in critical velocity gradient. Poor mixing in the premix tubes could result in hydrogen excursions from the bulk values across into the loop areas. Visual inspection of the Montebello test hardware suggests that such incipient flashback may have taken place and provides a degree of confidence in the validity of the loops. Also, the method of estimating flashback loops apears to be a reasonable one for preliminary design purposes.

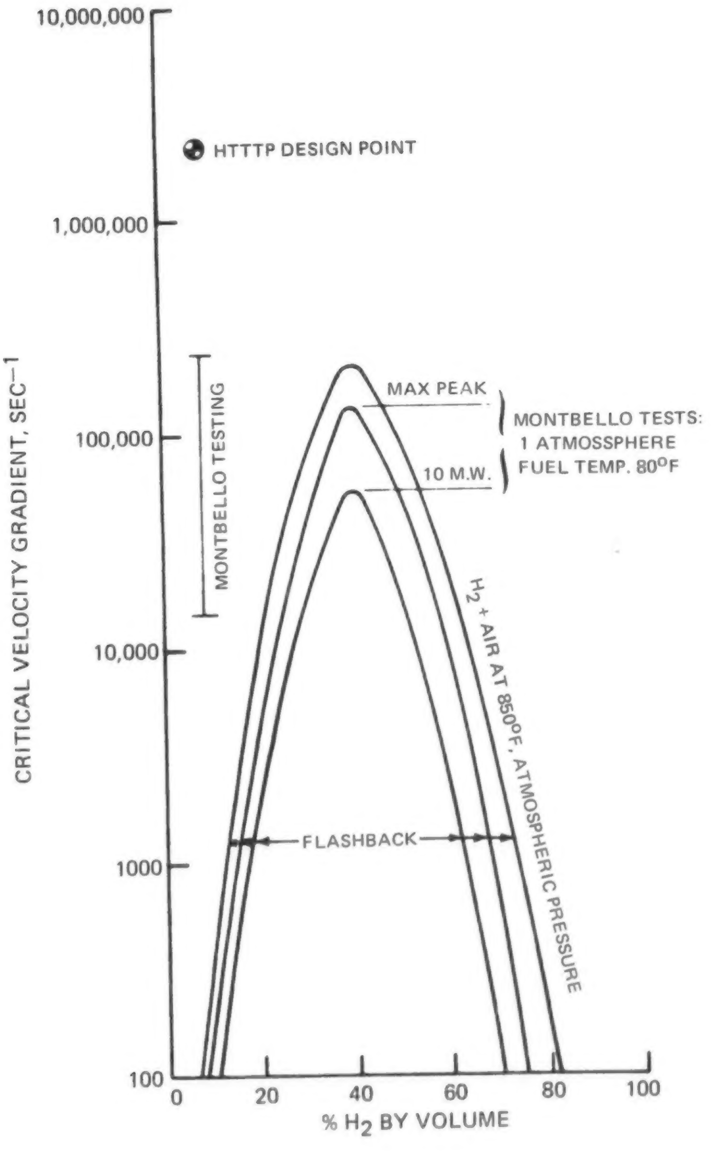
Also shown in Figure 9-17 is the operating point of the preliminary design burner. As indicated, it falls well outside the area of concern and gives a reasonable degree of confidence in the feasibility of the premix concept.

Staging--Premixed combustion systems have inherent limitations on the range of fuel-air ratios over which they operate. If the premixing is anywhere near complete, the mixture will burn at the bulk fuel-air ratio and not over a range of fuel-air ratios as with other systems. This means that combustion is not possible outside the bulk flammability limits of the mixture.

The proposed combustor is also different from most gas-turbine combustors in the manner in which the flame is stabilized. Most conventional combustors stabilize the flame front aerodynamically by means of swirling air and opposed air jets strategically placed to create rotating toroidal vortices into which the separate fuel supply is introduced. In the present case, the well-mixed fuel and total combustion air are introduced in the combustor as a large number of individual, unswirled jets containing all the reactants.

To hold the flame stable in such a system, it is necessary to use a mechanical flameholder which creates the necessary aerodynamic blockage. The requirements for such a flameholder are: it must provide stable combustion over the operating extremes expected in service; in conjunction with the remainder of the combustor, it must give good combustion efficiency at idle conditions; and it must achieve the above requirements without excessively long residence times which increase NO_x emissions.

ESTIMATED FLASHBACK LOOP–ESTIMATED FLASHBACK CHARACTERISTICS FOR THE PREMIX TUBE COMBUSTOR AND MONTEBELLO TEST ARE SHOWN



The type of dome design being considered provides aerodynamic blockage for flame stability through the faceplate material between individual premix tubes. Thus, the dome acts rather like a perforated plate flameholder, such as that described by Jamieson (Reference 9-29). The initial reactions on such flameholders take place in the multitude of little recirculation zones created between the individual mixture jets, and only later spread across the jets. These zones were not shown in Figure 9-15 for simplicity.

The stability characteristic must be defined for such flameholder when operating with low-Btu gas. The stability characteristic consists of a correlation of aerodynamic and geometric parameters of the flameholder, plotted against the mixture equivalence ratio so as to form a closed loop within which stable combustion is possible. United Technologies Corporation has demonstrated good success with premixed hydrocarbon/air flameholders using the Marquardt stability parameter (Reference 9-30), and it has been used again here. It is defined as:

$$\text{Stability Parameter} = \frac{V}{d} \cdot \frac{d}{d_e} \cdot \frac{P_o}{P} {\frac{1000}{T_o}}^{1.5} \text{X } 10^{-3} \qquad (9\text{-}8)$$

where, V = velocity of flow at flameholder
$\frac{de}{d}$ = effective dimension ratio, (≡ 1.6 for disks)
d = baffle dimension
P_o = reference pressure (≡ 1 atmos.)
P = static pressure in flow
T_o = mixture total temperature

For stable combustor operation it is necessary to always operate within the loops shown in Figure 9-18. These loops are much narrower than are those for more conventional gas turbine combustors which do not use premixing. The loops for premixed systems are usually not wide enough to satisfy most operating requirements. The design point mean equivalence ratio was selected as 0.78 on the basis of satisfying NO_x emissions requirements. This point is shown on Figure 9-18 inside the loop.

Once this point was decided, a totally fixed geometry system is forced to work along a smooth, continuous operating line which will eventually cross the lean limit of the stability loop at some part-power condition, causing a lean blowout of the combustor. To prevent lean blowout, it is necessary to use a variable geometry system. By making this part of the fuel system, it is possible to have the moving parts external to the engine flowpath. Variability of the fuel system is to be achieved by fueling the premix dome in discrete zones such that the functioning zone equivalence ratio is always kept inside the stability loop.

FIG. 9–18

CALCULATED STABILITY LOOP FOR PREMIXED CONCEPT AS APPLIED TO AN ANNULAR COMBUSTOR WITH 2600°F CET

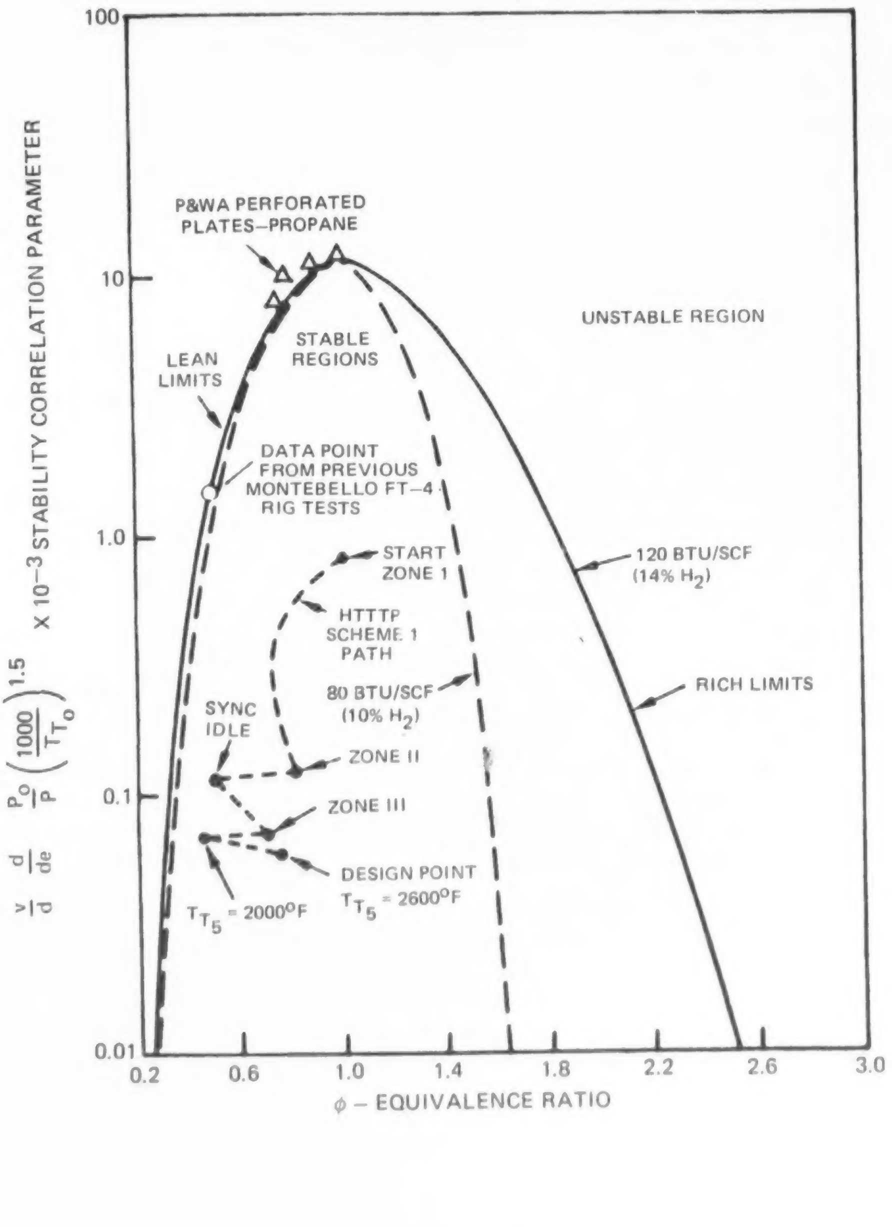


Figure 9-18, by way of illustration, displays the operating line for which a three-zone premix was done. At the design point all three zones are in operation. As power is reduced by reducing the fuel flow, the lean limit is approached until Zone III is shut down; and the total fuel flow at that point goes through Zones I and II only. This increases the equivalence ratio of these zones and moves the operating characteristic away from the lean limit. Further reduction in power causes the lean limit to be approached again until Zone II is shut down. This occurs just below synchronous idle speed. Start-up of the gas turbine is accomplished on Zone I at an equivalence ratio of unity for ease of starting. Zoning is achieved through means of internal baffling in the fuel manifold surrounding the premix tubes in the dome of the combustor.

Fuel Bound NO_x

While low-temperature cleanup systems achieve virtually complete removal of nitrogen compounds, possibly the most serious shortcoming of high-temperature cleanup systems is their inability to remove fuel-bound nitrogen. In conventional combustors with stoichiometric air or greater than stoichiometric air, almost all ammonia (at low, < 1% concentrations) has been shown to be converted to NO (Reference 9-11). Unless reliable and economical means can be developed to modify fuel composition in the gasifier, remove ammonia at high temperature or decompose it to nitrogen in the combustor, the viability of high-temperature cleanup is questionable.

Fortunately, with the support of the EPA, combustion methods have been demonstrated that would greatly reduce the amount of NO formed as a result of ammonia in the fuel gas. While it would be desirable to evaluate the merits of various methods of fuel bound NO_x control on a system level, such a study was outside the scope of this contract, but is recommended for further study.

Two rather similar schemes for reduction of fuel-bound NO_x are of interest. The first, reported in Reference 9-31 would utilize fuel rich combustion with long residence times to decompose the undesirable species. The effect of residence time and equivalence ratio is shown in Figure 9-19 and the conceptual design of the combustor in Figure 9-20. The secondary air would be added in a rapid mixing zone to minimize formation of thermal NO_x. Because of the long residence time, burner design would be nonconventional and quite large as compared to current designs. While such a burner could prove costly when used in integrated systems such as proposed here, a combined burner-pressurized boiler could be an interesting alternative.

A variation of the above concept, being studied under EPA Contract No. 68-02-2136 at the Pratt & Whitney Aircraft Division of UTC, would result in a more conventional size burner. The concept (Reference 9-32) also uses a rich burn followed by a very short residence time in the first zone. Results of that work are shown in Figure 9-21. If all nitrogen in that fuel were to appear in the stack as NO, its concentration would be 191 ppm.

FIG. 9–19

RESIDUAL NITROGEN SPECIES AS A FUNCTION OF EQUIVALENCE RATIO AND RESIDENCE TIME

NH_3^o = 4000 PPM IN FUEL
T_{FUEL} = 18 1088K
T_{AIR} = 810K
PRESSURE = 10 ATM
RESIDUAL NITROGEN SPECIES = Σ (NO, NH_3 AND HCN)

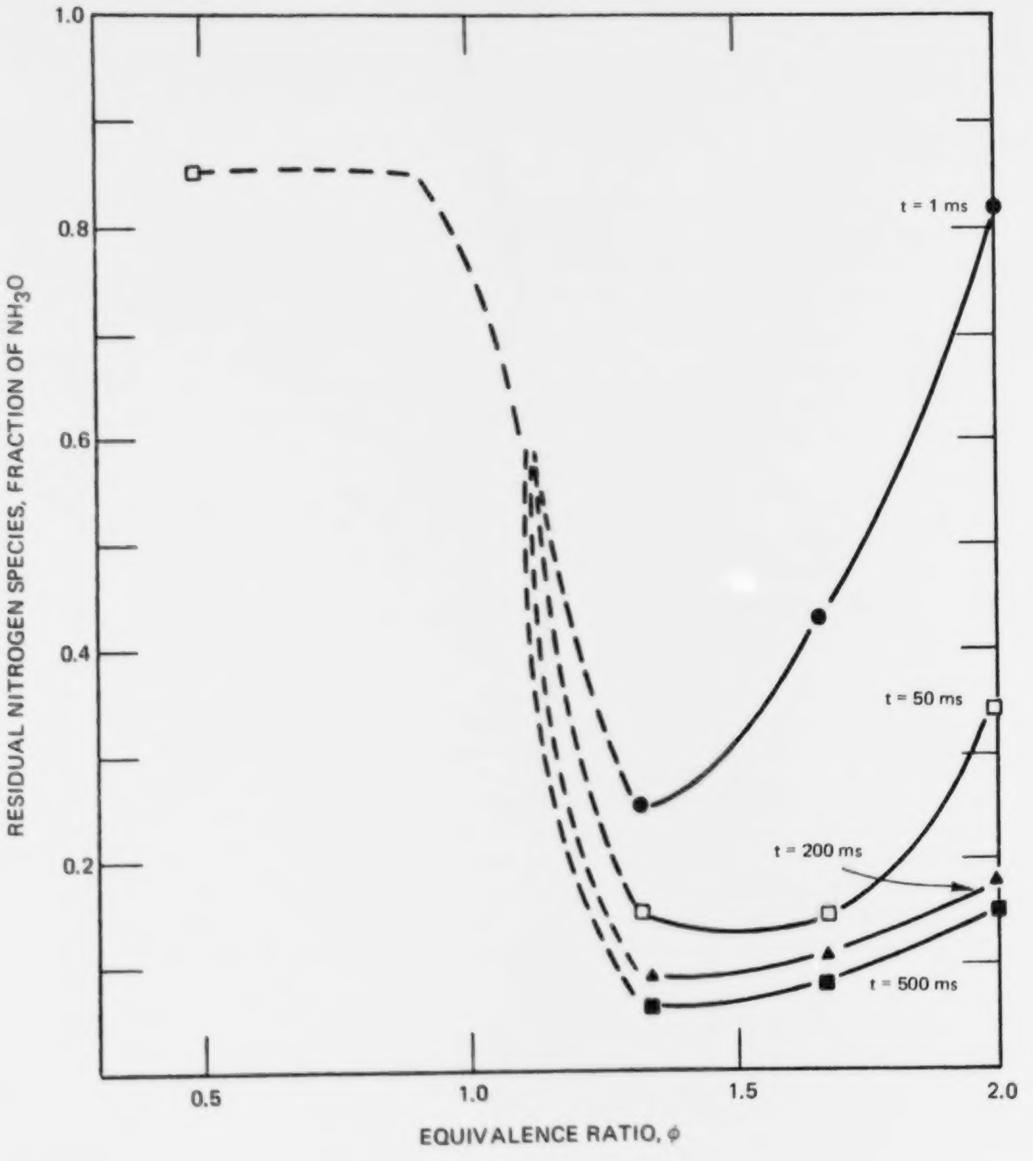


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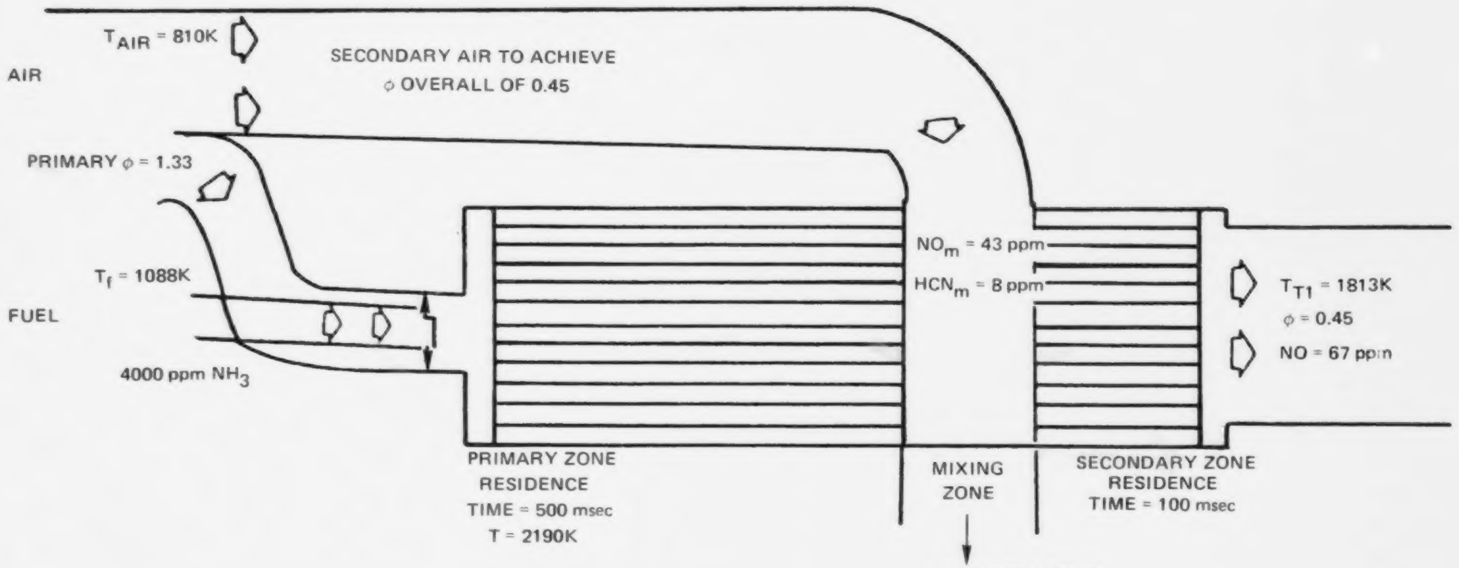
FIG. 9–20

OPTIMIZATION OF THE INITIAL DESIGN CONCEPT


AIR
T_{AIR} = 810K
SECONDARY AIR TO ACHIEVE
ϕ OVERALL OF 0.45
PRIMARY ϕ = 1.33
T_f = 1088K
FUEL
4000 ppm NH_3
NO_m = 43 ppm
HCN_m = 8 ppm
T_{T1} = 1813K
ϕ = 0.45
NO = 67 ppm
PRIMARY ZONE
RESIDENCE
TIME = 500 msec
T = 2190K
MIXING
ZONE
SECONDARY ZONE
RESIDENCE
TIME = 100 msec

INSTANTANEOUS

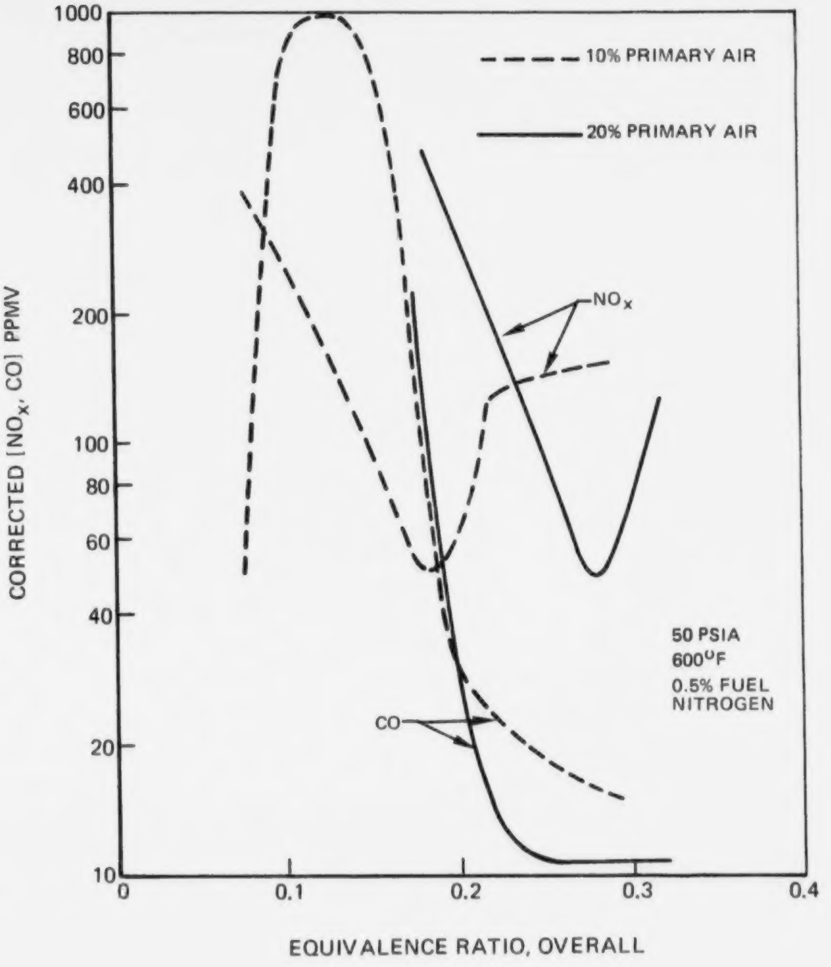
NOTE: NO_m AND HCN_m INCLUDE
SECONDARY DILUTION EFFECT

78–02–157–1

FIG. 9–21

RICH BURNER STAGING CHARACTERISTICS

RICH BURN/QUICK QUENCH CANCEPT



78–02–157–3

Thus, the observed reduction to 50 ppm is significant. Also, for the case with 20% primary air, CO emission levels are quite low. Apparently it is possible to move the position of minimum NO formation while maintaining a relatively constant CO characteristic.

PARTICULATES

The effluent from each gasifier will contain varying amounts of entrained particulate matter. To date, no quantitative assessment has been made of particulate loadings in gases produced by the specified coal gasification processes. The Koppers-Totzek gasifier effluent has been characterized as having an entrained particulate concentration of 11.57 gr/scf (Reference 9-33). Westinghouse has prepared a quantitative approximation of raw gas particulate loading and size distribution for fluidized bed gasification. The projection is based on fluidized bed combustion and a theoretical model developed by Kunii and Levenspiel (References 9-33, 9-34). The Westinghouse projections and the particulate loads estimated for the gasifiers studied here are shown in Table 9-8. The molten salt bed Kellogg gasifier removes particulates before they exit the gasifier. For this reason a lower particulate loading, nearly all entrained sodium salts, is expected.

A summary of removal system performance developed by Westinghouse (Reference 9-34) is shown in Figure 9-22. Since the gas turbine is believed to be tolerant to particles of less than 2 micron diameter, the venturi type scrubber appears ideally suited for this duty. It is also apparent that those concepts amenable to high-temperature operation have significantly reduced removal capabilities in the size range of interest.

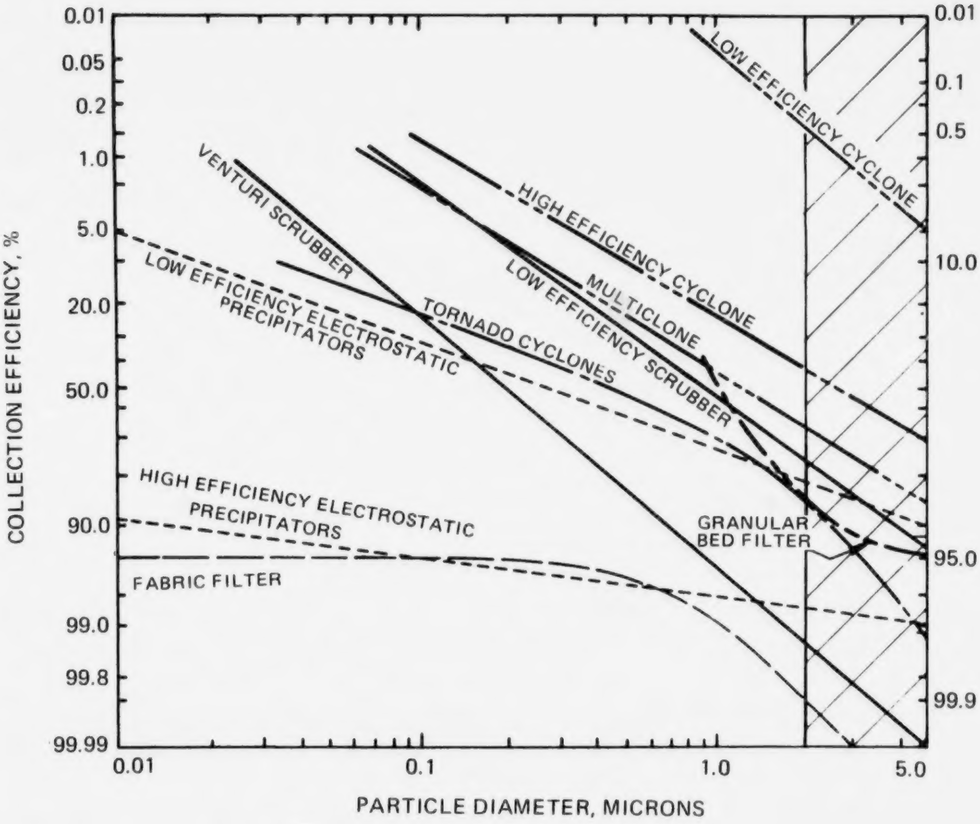
For the high-temperature systems studies here, use of a granular bed filter was assumed. Particulate removal down to .01 to .02 grain/SCF is estimated (Reference 9-35). Other candidates include a woven metal cloth manufactured by the Brunswick Corporation and the panel bed sand filter being developed at City College, New York (References 9-36, 9-37, and 9-38). Performance of the panel bed filter is reported to be as high as 99.99% with no particles larger than 5 micron penetrating the filter. For the purposes of this study, it is important that a representative cost for particulate removal be used. Data obtained from References 9-35 , 9-38 and 9-39 all show installed costs for the various concepts to be in the range of \$80 to \$100 per actual CFM for units of the size used here.

TABLE 9-8

PARTICULATE LOADING IN GASIFIER PRODUCT GASES

Process Configuration	Projected Dust Load	Particle Size Distribution
Westinghouse (Estimate for Fluidized-Bed Gasification)	10 to 30 gr/SCF	10 to 25% < 10 μ 5 to 15% < 5 μ
BCR-Air (Entrained Bed)	15 to 45 gr/SCF	15 to 35% < 10 μ 10 to 20% < 5 μ
BCR-Oxygen (Entrained Bed)	15 to 45 gr/SCF	15 to 35% < 10 μ 10 to 20% < 5 μ
IGT-Air (Fluidized Bed)	5 to 25 gr/SCF	10 to 25% < 10 μ 5 to 15% < 5 μ
Kellogg-Air (Molten Bed)	0.1 to 1 gr/SCF	10 to 30% < 10 μ 5 to 20% < 5 μ

EXTRAPOLATED FRACTIONAL EFFICIENCY OF PARTICULATE REMOVAL DEVICE



78–02–157–6

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SECTION 10

OTHER ENVIRONMENTAL INTRUSIONS

INTRODUCTION

The previous section has dealt with the major air emissions from the integrated power plant. With the exception of thermal NO_x, all the potential pollutants arose because of the nature of the coal feedstock. From the time that the coal enters the plant boundary to the time that solid wastes exit, there is the potential to cause some type of environmental intrusion. Except for some chemicals introduced for water treatment, the original coal feed is the source for all the major air, water, solid waste and minor and trace element emissions.

The organic portion of coal is largely composed of polycyclic aromatic structures and, in addition to carbon and hydrogen, contains significant quantities of organic oxygen, nitrogen, and sulfur, residual biological compounds, and other organic compounds. Functional groupings include methoxyl, hydroxyl, carbonyl, and carboxyl. The form into which organic sulfur and nitrogen combine with other elements or compounds in the coal is not known, but it is presumed to be largely or entirely in heterocyclic aromatic rings. Coal mineral matter contains trace to appreciable quantities of practically all the elements in the periodic table.

The proximate and trace-elements analysis of coals may vary considerably from mine to mine, or even from seam to seam (Reference 10-1). The proximate analysis for coal lists only moisture, volatile matter, fixed carbon, and ash. An ultimate analysis is only slightly more definitive; a complete ultimate analysis for typical Illinois No. 6 coal is listed in Table 10-1 from Reference 10-2. Typical modes of occurrence of trace and minor elements in coal are given in Table 10-2 (Reference 10-3).

A general procedure that considers concentration and toxicity of trace elements is described for choosing those most likely to be of environmental concern. Using that procedure, the trace elements arsenic, beryllium, boron, cadmium, chromium, lead, mercury, nickel and vanadium have been selected for

TABLE 10-1

TYPICAL COMPOSITION OF ILLINOIS NO. 6 COAL

Proximate Analysis, As Received	Wt-%
Moisture	13.3
Volatile Matter	33.0
Ash	8.8
Fixed Carbon	44.9
Total	100.0
Ultimate Analysis, Dry Basis	
Carbon	70.1
Hydrogen	4.88
Sulfur	3.74
Nitrogen	1.07
Oxygen (by difference)	10.11
Ash	10.10
Minor Elements	
Silicon	2.0
Iron	1.4
Calcium	0.35
Chlorine	0.23
Potassium	0.17
Sodium	0.14
Trace Elements	**ppm**
Titanium	700
Magnesium	570
Boron	200
Fluorine	61
Zinc	49
Manganese	48
Strontium	37
Zirconium	35
Lithium	33
Barium	31
Arsenic	24
Copper	19
Vanadium	17
Chromium	15
Nickel	1
Selenium	13
Lead	11
Tellurium	8.1
Molybdenum	7.0
Germanium	4.3
Cobalt	3.6
Tin	2.0
Antimony	1.1
Bismuth	1.1
Beryllium	1.0
Cadmium	0.89
Samarium	0.74
Ytterbium	0.56
Mercury	0.12
Silver	0.10

TABLE 10-2

TYPICAL FORM IN WHICH TRACE AND MINOR ELEMENTS OCCUR IN COAL

Element	Form	Element	Form
Sb	Sulfide	Mo	Sulfide
As	Oxide, sulfide	Ni	Sulfide
Ba	Carbonate, sulfate with Ca	N	ORO
Be	ORO[*]	K	KCl, carbonate
Bi	Sulfide	Sm	SQ
B	ORO, borate	Sc	Oxide
Cd	Sulfide	Se	PORO, sulfide, iron selenides
Ca	Oxide, carbonate, sulfate	Si	Oxide, SQ
Cl	PORO[+] sodium chloride	Ag	Element, sulfide, SQ
Cr	PORO, oxide	Na	PORO, carbonate
Co	PORO, sulfide	Sr	PORO, with Ca
Cu	$CuFeS_2$, sulfide	S	PORO, sulfides, sulfates
F	CaF_2	Te	Iron tellurides
Ge	PORO, carbonate	Th	SQ
Fe	Carbonate, sulfide, oxide	Sn	Carbonate, sulfide
Pb	Sulfide	Ti	PORO, SQ
Li	SQ[**]	V	PORO
Mg	PORO, carbonate, SQ	Yb	SQ
Mn	Carbonate in $CaCO_3$, SQ	Zn	Sulfide
Hg	PORO, elemental, sulfide	Zr	Oxide, SQ

* ORO - Organic occurrence

+ PORO - Partial organic occurrence

** SQ - Silicates, clay, quartz

discussion of their fate in the integrated plant processes. All these elements except boron and vanadium are included on EPA's list of criteria pollutants. Because of the potentially corrosive activity of boron and vanadium salts in the hot section of the gas turbine, these elements have also been selected for further study.

The following presents a general discussion of the sources of pollution in an integrated plant and describes, wherever possible, methods of controlling the air, water or solid waste emissions. The trace elements, in particular those chosen for examples, are also discussed.

WASTES FROM COAL PREPARATION AND HANDLING

Figure 10-1 is a schematic block diagram of coal handling and preparation facilities. Run-of-mine (ROM) coal will be received from the mine by trucks. The amount of coal stored at a commercial-sized, combined-cycle power plant of 1,000-MW capacity will depend upon the availability of coal from the mine, the location of the plant, and storage space available at the plant. Coal will be received on a continuous basis and stored in an active coal pile. It is a common practice to stockpile enough coal for a three to fifteen weeks supply. This coal will be stored in an inactive pile and will be utilized in case of unusual circumstances, such as unavailability of coal on a continuous basis due to bad weather or other uncontrollable conditions.

Size Reduction

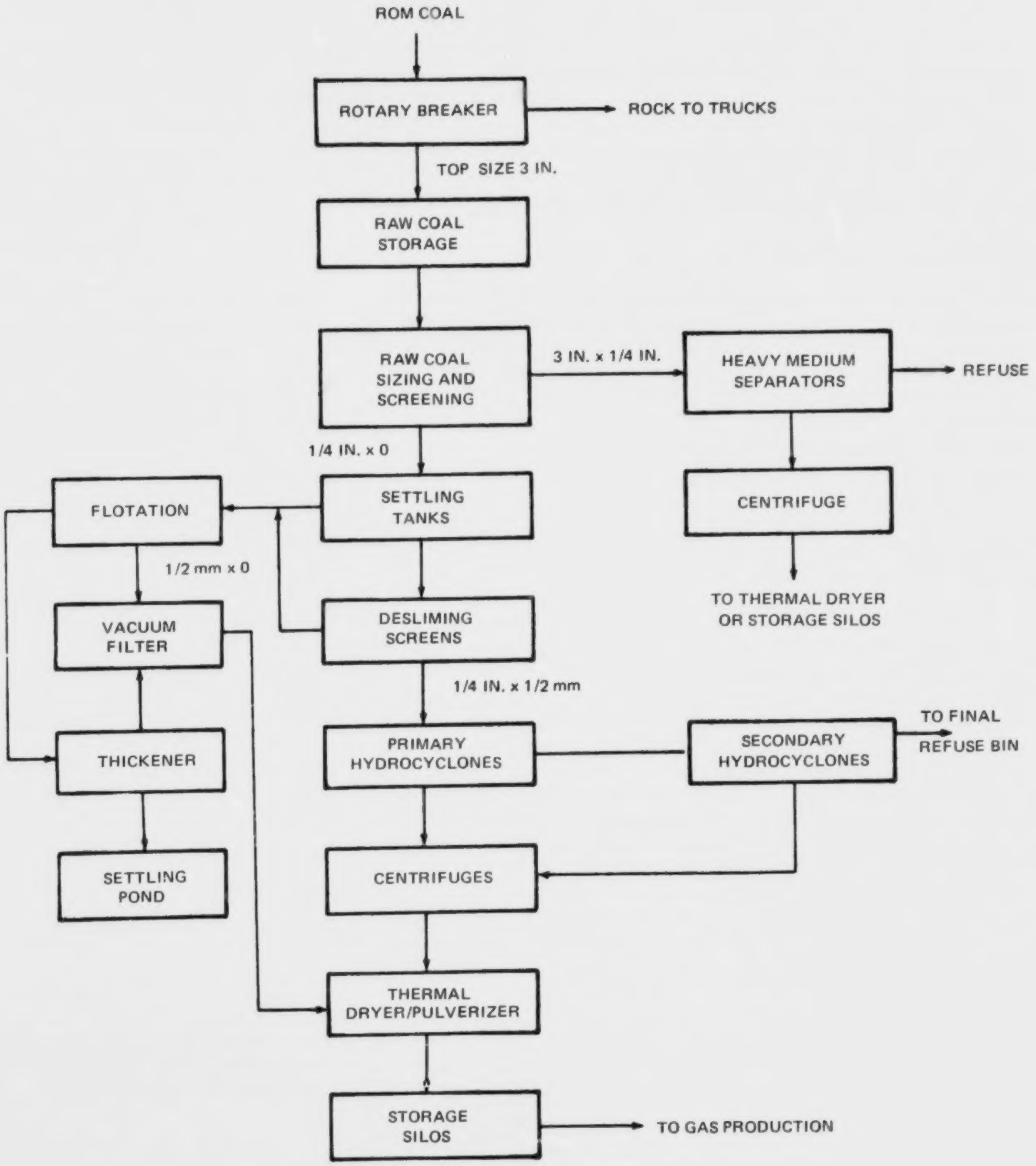
Coal size reduction is done in stages. Primary crushers, either roll-type or rotary screen-type, are employed to reduce the raw coal to a top size of 3 inches. Coal is reduced further with secondary crushers and screening crushers. Secondary crushers will reduce coal to a top size of from one-and-one-half to one-and-three-fourths inches, screening crushers will reduce coal to a top size of from 3/8 to 1 inch. Different types of crushers are used for coal cleaning; from an environmental point of view, crushers producing lesser amounts of fines are desirable. The commonly used crushers are hammer-mill and ring crushers, single-roll crusher, double-roll crusher, rotary breaker, and pick breaker.

Screening

Screening is used for sizing coal for further processing. The crushed coal is sized into various fractions: 3" x 1/4", which is further processed in heavy medium separators; and 1/4" x 0 which is further sized by desliming screens into two different sizes, 1/4" x 1/2 mm and 1/2 mm x 0. Coal particles may be screened wet or dry. The various types of screening equipment used include gravity-bar screens (or grizzlies), revolving screens, shaking screens, vibrating screens, and sieve bends. Wet screening produces large quantities of contaminated water whereas dust is generated by dry screening.

FIG.10–1

COAL PREPARATION AND HANDLING


ROM COAL
ROTARY BREAKER
ROCK TO TRUCKS
TOP SIZE 3 IN.
RAW COAL STORAGE
RAW COAL SIZING AND SCREENING
3 IN. x 1/4 IN.
HEAVY MEDIUM SEPARATORS
REFUSE
CENTRIFUGE
TO THERMAL DRYER OR STORAGE SILOS
1/4 IN. x 0
FLOTATION
SETTLING TANKS
1/2 mm x 0
VACUUM FILTER
DESLIMING SCREENS
1/4 IN. x 1/2 mm
THICKENER
PRIMARY HYDROCYCLONES
SECONDARY HYDROCYCLONES
TO FINAL REFUSE BIN
SETTLING POND
CENTRIFUGES
THERMAL DRYER/PULVERIZER
STORAGE SILOS
TO GAS PRODUCTION

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Thermal Drying

Moisture content in the feed coal acceptable to different gasifiers will vary, depending upon the gasifier configuration. During coal sizing and screening, the water content of coal will increase to a level as high as 10 to 15 percent. Generally, the moisture content of coal is reduced to 2 to 6 percent by drying with hot gases in thermal dryers. The hot gases are usually the gaseous combustion effluent from a burner or air heater. After being used for drying the wet coal, this gas is vented to the atmosphere from an elevated vent. Temperature control is essential in drying to prevent spontaneous combustion when the drying medium contains high levels of oxygen. The types of thermal dryers available are rotary dryers, screen-type dryers, fluidized-bed dryers, cascade dryers, and multilouver dryers.

Air Pollution and Control

Atmospheric discharges from coal preparation units appear to be the largest single source of particulate emissions from the combined-cycle plant if coal is prepared on site. The major sources of particulate emissions, including fugitive dust, are shown in Figure 10-2. Dust is composed of coal particles, typically 1 to 100 microns in size, similar to the parent coal in composition. Fugitive dust is generated from coal unloading, crushing, screening, and conveying operations. More coal dust will be generated for gasifiers requiring smaller particle sizes. Pneumatic conveying of coal also will generate more dust and better control will be necessary. Inactive coal piles do not cause a major dust problem but active coal piles could have higher than usual dust emissions under high wind conditions. Coal stored in silos will produce no dust emission.

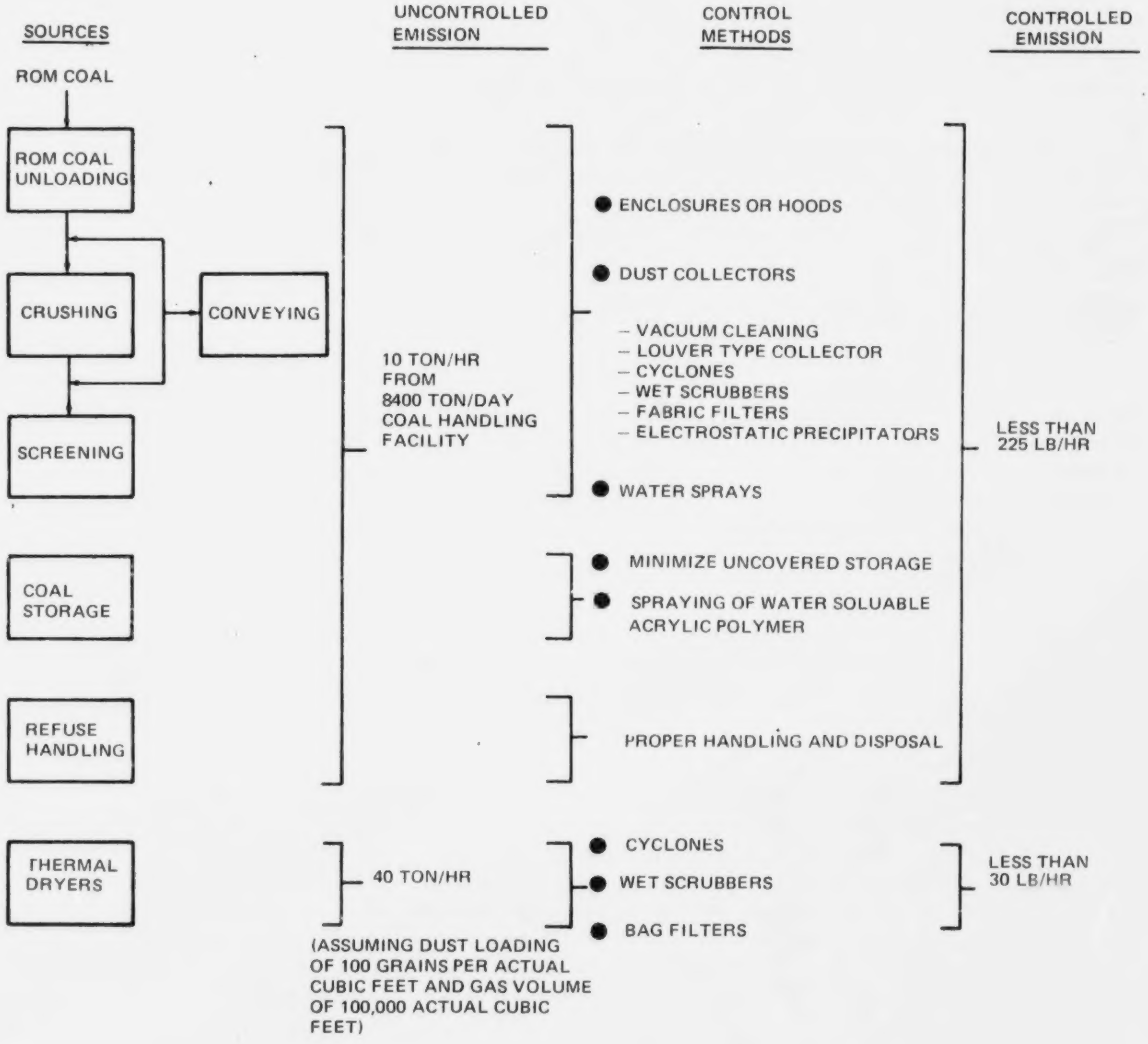
Coal dryers in the combined-cycle power system will utilize the stack gas from the heat recovery boiler. This gas is available at a temperature of about 300F. Major components present in the stack gas are carbon dioxide, nitrogen, oxygen and water with small amounts of NO_x and SO_2 that are generated during the combustion of the low-Btu gas in the turbine burner. The exhaust gas from the dryer will include, in addition to the components above, coal particulates, water and trace amounts of volatile organics that may be present in the coal. The exhaust gas leaves the dryer at a temperature of about 220F. The coal is heated to a temperature of about 200F. Both BCR air- and oxygen-blown processes use part of the stack gas for coal drying. For a 1,000-MW plant, a flow rate of about 80,400 moles/hr is required. The U-Gas gasifier does not require coal drying. In general, vent gases from the dryer will have a higher moisture content than the stack gas.

Particulate Control Methods

The particulate control methods applicable to different sources of particulate dust generation also are shown in Figure 10-2. Crushing, conveying, pulverizing, loading and unloading equipment can be operated with enclosures or hoods. Water sprays can be used to prevent dispersion of dust

FIG. 10–2

PARTICULATES EMISSIONS FROM COAL PREPARATION AND HANDLING


SOURCES
UNCONTROLLED EMISSION
CONTROL METHODS
CONTROLLED EMISSION
ROM COAL
ROM COAL UNLOADING
CRUSHING
CONVEYING
SCREENING
COAL STORAGE
REFUSE HANDLING
THERMAL DRYERS
10 TON/HR FROM 8400 TON/DAY COAL HANDLING FACILITY
● ENCLOSURES OR HOODS
● DUST COLLECTORS
– VACUUM CLEANING
– LOUVER TYPE COLLECTOR
– CYCLONES
– WET SCRUBBERS
– FABRIC FILTERS
– ELECTROSTATIC PRECIPITATORS
● WATER SPRAYS
● MINIMIZE UNCOVERED STORAGE
● SPRAYING OF WATER SOLUABLE ACRYLIC POLYMER
PROPER HANDLING AND DISPOSAL
LESS THAN 225 LB/HR
40 TON/HR
● CYCLONES
● WET SCRUBBERS
● BAG FILTERS
LESS THAN 30 LB/HR
(ASSUMING DUST LOADING OF 100 GRAINS PER ACTUAL CUBIC FEET AND GAS VOLUME OF 100,000 ACTUAL CUBIC FEET)

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particles at loading and unloading points, at transfer and discharge points from crushing and screening, and from the coal storage area. Outdoor coal storage piles should be covered with coatings, such as acrylic polymer, to provide protection from wind and rain.

Application of suitable dust collection devices will also help reduce the particulate emissions. Vacuum-cleaning systems can effectively minimize the particulate dispersion around the plant. The louver-type collectors can operate at 99 percent efficiency when collecting plus 100-mesh dust and are suitable for collecting heavy dust. Primary cyclones can effectively collect particles larger than 200 mesh. Secondary cyclones in series with primary cyclones can collect all the plus 10-micron particles. Wet scrubbers can remove submicron particles, but consume more energy. Fabric filters can operate at a 99 percent efficiency, but their application is limited to temperatures below 500F. Electrostatic precipitators can remove submicron particles.

The particulate loading rate from thermal dryers, if uncontrolled, can be as high as 100 to 300 grains per actual cubic foot. Such a high emission rate can be controlled by operating primary and secondary cyclones in the raw gas transfer line. A wet scrubber can also be used in place of secondary cyclones.

Water Pollution and Control

Two major sources of wastewater from the coal preparation and handling area are the water generated from coal processing units such as wet screening and the runoff water from raw coal storage and coal waste deposits. Water from the processing units contains fine and colloidal particles, dissolved mineral matter, and salts, all of which were present in coal. Chemical agents added to this water for solids removal by flotation and flocculation are also present. Approximately 2600 gpm of wastewater will be generated from processing 350 TPH (8400 TPD) of raw coal. However, most of it will be treated and recycled for reuse.

The second source of wastewater, runoff water from coal storage and waste deposits, is contaminated with high concentrations of sulfate and metal ions. The amount of runoff water will depend on the rate of precipitation at the site.

Coal Pile Runoff Water Characteristics--

Coal pile runoff is the drainage from the coal storage area usually resulting from rain. Runoff may cause pollution if allowed to enter waterways or to seep into aquifers. The nature of coal pile runoff depends on the type of coal used. Generally, there are two types of coal pile runoff. The first is neutral or slightly alkaline, contains ferrous ion, and originates from coal containing large amounts of alkaline materials or small amounts of pyrite. The second kind of runoff is highly acidic, contains large amounts of dissolved iron and aluminum, and is produced from pyrite-rich coal.

Pyrite is oxidized by atmospheric oxygen and hydrolyzed to form ferrous sulfate and sulfuric acid according to the reaction

$$2FeS_2 + 7O_2 + 2H_2O \rightarrow 2FeSO_4 + 2H_2SO_4. \qquad (10\text{-}1)$$

Additional sulfuric acid may be formed if the ferrous ion is further oxidized to the ferric state. When rain falls on coal piles, the acid is washed out and eventually becomes part of the coal pile drainage. At low pH, metals such as aluminum, copper, manganese, zinc, and others are also dissolved, further degrading the water.

Coal pile runoff water characteristics can vary widely, depending on the coal characteristics, weather conditions, and the length of contact between coal and water. The amount of rainfall that percolates through coal storage piles is a function of the area of the pile and the rainfall. The area of the coal storage pile for a 1000-MW electric plant may range from 12 to 60 acres, depending on how high the coal is piled and the number of operating days the plant has in storage.

Characteristics of coal pile runoff at selected steam-electric power plants are given in Table 10-3 (Reference 10-4). The coal types are unknown. This water is treated along with other waste streams from the integrated plant.

Water Handling and Treatment--

Water with less than 5 percent solid matter, neutral pH, low conductivity, and low bicarbonate can be used in coal preparation plants. The wastewater from the coal preparation units can be handled and treated to this quality by a closed water circuit consisting of thickeners, cyclones, filters, and/or solid bowl centrifuges. Solids are separated by sedimentation. Flocculation increases the size of particles, thereby aiding the sedimentation process. Various flocculation agents are employed for treatment, among which starches are probably those most commonly used.

Thickeners are used to remove solids from contaminated water containing less than 10 percent solids. The overflow from the thickener may contain less than 1 percent solids of very fine size that cannot settle. The underflow is 60 percent solids. Cyclones are also used to remove suspended solids and high ash slimes from the water. Water may be treated in hydrocyclones before it is routed to the thickeners. The spinning force in hydrocyclones separates the solids by forcing them to the outer edge and discharging them as a thick slurry at the bottom of the cone. Fine particles and ash slimes are discharged at the top. Typical performance of a 3-inch diameter hydrocyclone shows that 60 mesh by 0 material containing 10 percent solids is separated into an overflow of 4 percent solids and an underflow of 35 percent solids. The underflow from hydrocyclones or thickeners is routed to settling ponds or further concentrated in centrifuges or vacuum filters, resulting in coarse solid waste, most of which can be recycled for reuse. A simple flow scheme for water handling is shown in Figure 10-3.

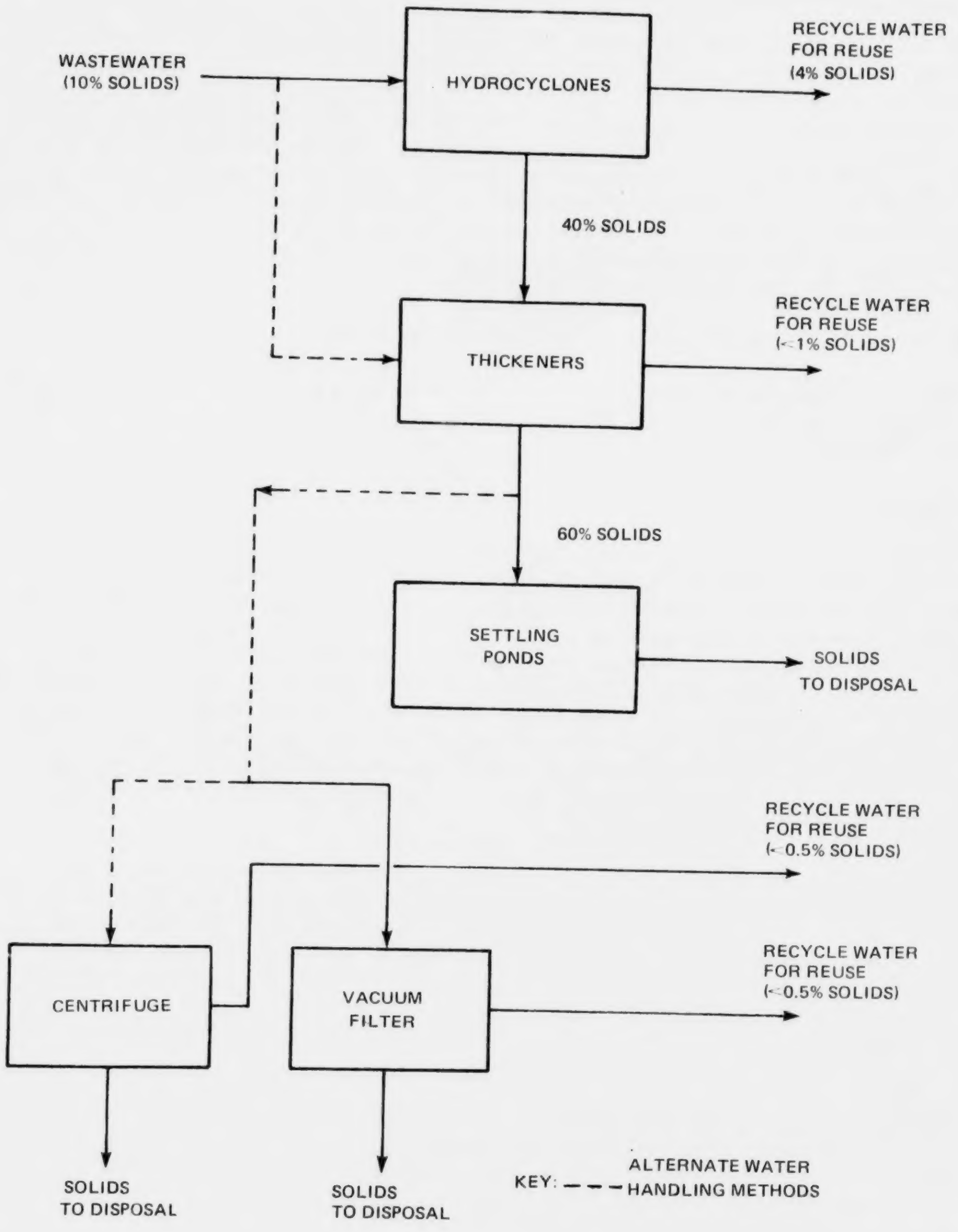
TABLE 10-3

COAL-PILE RUNOFF ANALYSIS AT SELECTED PLANTS
(mg/l)

Plant	A	B	C	D	E	F	G
Alkalinity	6	0	-	-	14.32	36.41	-
Total solids	1,330	9,999	-	-	-	-	6,000
TDS	720	7,743	-	28,970	-	-	5,800
TSS	610	22	-	100	-	-	200
Ammonia	0	1.77	-	-	-	-	1.35
Nitrate	0.3	1.9	-	-	-	-	1.8
Phosphorous	-	1.2	-	-	-	-	-
Turbidity	505	-	-	-	2.77	6.13	-
Acidity	-	-	-	21,700	10.25	8.84	-
Total hardness	130	1,109	-	-	-	-	1,851
Sulfate	525	5,231	6,837	19,000	-	-	861
Chloride	3.6	481	-	-	-	-	-
Aluminum	-	-	-	1,200	-	-	-
Chromium	0	0.37	-	15.7	-	-	0.05
Copper	1.6	-	-	1.8	-	-	-
Iron	0.168	-	0.368	4,700	1.05	0.9	0.06
Magnesium	0	89	-	-	-	-	17.4
Zinc	1.6	2.43	-	12.5	-	-	0.006
Sodium	1,260	160	-	-	-	-	-
pH	2.8	3	2.7	2.1	6.6	6.6	4.4

FIG. 10–3

WATER HANDLING FOR COAL PREPARATION


WASTEWATER
(10% SOLIDS)
HYDROCYCLONES
RECYCLE WATER
FOR REUSE
(4% SOLIDS)
40% SOLIDS
THICKENERS
RECYCLE WATER
FOR REUSE
(<1% SOLIDS)
60% SOLIDS
SETTLING
PONDS
SOLIDS
TO DISPOSAL
RECYCLE WATER
FOR REUSE
(<0.5% SOLIDS)
RECYCLE WATER
FOR REUSE
(<0.5% SOLIDS)
CENTRIFUGE
VACUUM
FILTER
SOLIDS
TO DISPOSAL
SOLIDS
TO DISPOSAL
KEY: – – – ALTERNATE WATER
HANDLING METHODS

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Wastewater resulting from raw coal piles or coal waste piles can be treated by neutralization. Acid or alkali is added to neutralize alkaline or acidic wastewater. Common alkalies used are lime, limestone, or soda ash. Sulfuric acid is used to neutralize alkaline water.

Solid Waste and Disposal

It is estimated that about 25 percent of the raw coal mined is disposed of as waste. For a coal conversion plant processing 8400 TPD of prepared coal, the coal refuse will amount to 2100 TPD. The coal refuse consists of waste coal, slate, carbonaceous and pyritic shales, and clay asociated with a coal seam. There are two basic types of coal refuse, coarse and fine. The coarse refuse is separated mainly by crushing the lump coal to a smaller size; the fine refuse is mainly separated from the thickeners as underflow and impounded into the settling ponds. The coarse refuse is normally disposed of in an embankment. Further discussion of the disposal of these wastes is presented in the discussion for the integrated plant.

WASTES FROM GASIFICATION, CLEANUP AND GAS UTILIZATION

The major air emissions resulting from the preparation and combustion of the fuel gas have been identified in Section 9. Thus, the emphasis here will be upon potential water borne and solid wastes.

Waste Water Sources

Water requirements for a coal conversion plant are high. Variation of water requirements results from a combination of process requirements, plus cooling and wastewater streams subjected to varying degrees of conservation and reuse. Low-Btu coal gasification requires about 1 to 3 pounds of water per pound of feed coal. Low-Btu gasification combined with cleanup and a COGAS power cycle will need an additional 2 to 4 pounds of water per pound of feed coal. This water is used in these processes for different objectives. Steam is required for gasification, water is used for scrubbing the raw gases from the gasifier, a large amount of steam is generated in the boilers for use in steam turbines, and circulating cooling water removes excess heat generated in the process.

Part of the water entering the process comes out as wastewater. A portion of the steam fed to the gasifier remains unreacted and exits with raw gas. It is condensed in knock-out drums and combined with the sour water stream. Raw gas from the gasifier is scrubbed with water to absorb ammonia and to lower further the raw gas temperature. The scrubbing water also dissolves hydrogen sulfide and carbon dioxide, and phenols, cyanides, and organics that may be present. This constitutes the sour water stream and is the dirtiest water generated in the process.

Regeneration wastes are produced during resins regeneration of the demineralizers used to deionize boiler feedwater. Cooling tower blowdown is another source of wastewater. Raw water treatment wastes and boiler cleaning wastes are also generated from the power system. These sources are shown in Figure 10-4. Coal pile drainage and yard runoff wastes will also need treatment before they are discharged. Figure 10-5 shows the sources of wastewater from the integrated plant and the anticipated treatment. The characteristics of these wastewaters, the applicable treatment methods, and possible reuse of the treated water are described in the following sections.

Wastewater Characteristics

The characteristics of process sour water, coal pile drainage, boiler and cooling tower blowdown, water treatment wastes, equipment cleaning wastes, and floor and yard drains have been identified. These categories include all types of wastewaters generated by the facility.

Sour Water Characteristics--

Sour water is the dirtiest wastewater generated in the process. A portion of it originates from the use of steam within the gasifiers. The subsequent condensation of the unused steam usually occurs simultaneously with the condensation of hydrocarbon liquids in the offgas from fixed and fluidized bed gasifiers. The vapor phase contains hydrogen sulfide, ammonia, carbon dioxide, and volatile cyanides as major impurities. The water condensed in the knock-out drum contains these gases. The gaseous mixture is further cooled by water scrubbing. Part of the hydrogen sulfide, carbon dioxide, and ammonia dissolve in the scrubbing water. Small amounts of phenols and other organics may also be present in the case of the fluidized bed gasifier. However, the Institute of Gas Technology claims that phenols and organics are not produced in their fluidized bed gasifier.

Sour water characteristics will depend on the composition of the raw gas, the scrubber design, and its operating conditions. Sour water characteristics are not available for the processes considered in this work; however, information from other processes has been used to determine the sour water characteristics. A by-product water analysis from Synthane gasification of various coals is given in Table 10-4 from Reference 10-5. Approximately 60 percent of the nitrogen in coal is converted to ammonia. The concentration of cyanide is notably small (0.6 mg/l or lower). Thiocyanate levels are also low, compared to coke plant weak ammonia liquor. The variation in phenol concentration for different coal feeds is wide, from 1,700 to 6,000 mg/l. The entrained bed BCR gasifier does not produce phenols; thus they will be absent in the wastewater generated in the scrubbing operation of this configuration. Sour water may contain as much as 2 to 5 percent by weight of ammonia and 2 to 6 percent by weight of hydrogen sulfide. Suspended char particles also may be present in the wastewater from the scrubbing operation. Table 10-5 shows the composition of gasification wastewater as estimated by the Bechtel Corporation (Reference 10-6). This wastewater composition was compiled from a number of sources including process design data and analyses of other gasification liquors. The sour water characteristics of Table 10-4 and 10-5 are not used in the present study; but these characteristics indicate the types of compounds that may be present in the wastewater.

FIG. 10–4

SOURCES OF WASTEWATER IN A COGAS POWER SYSTEM

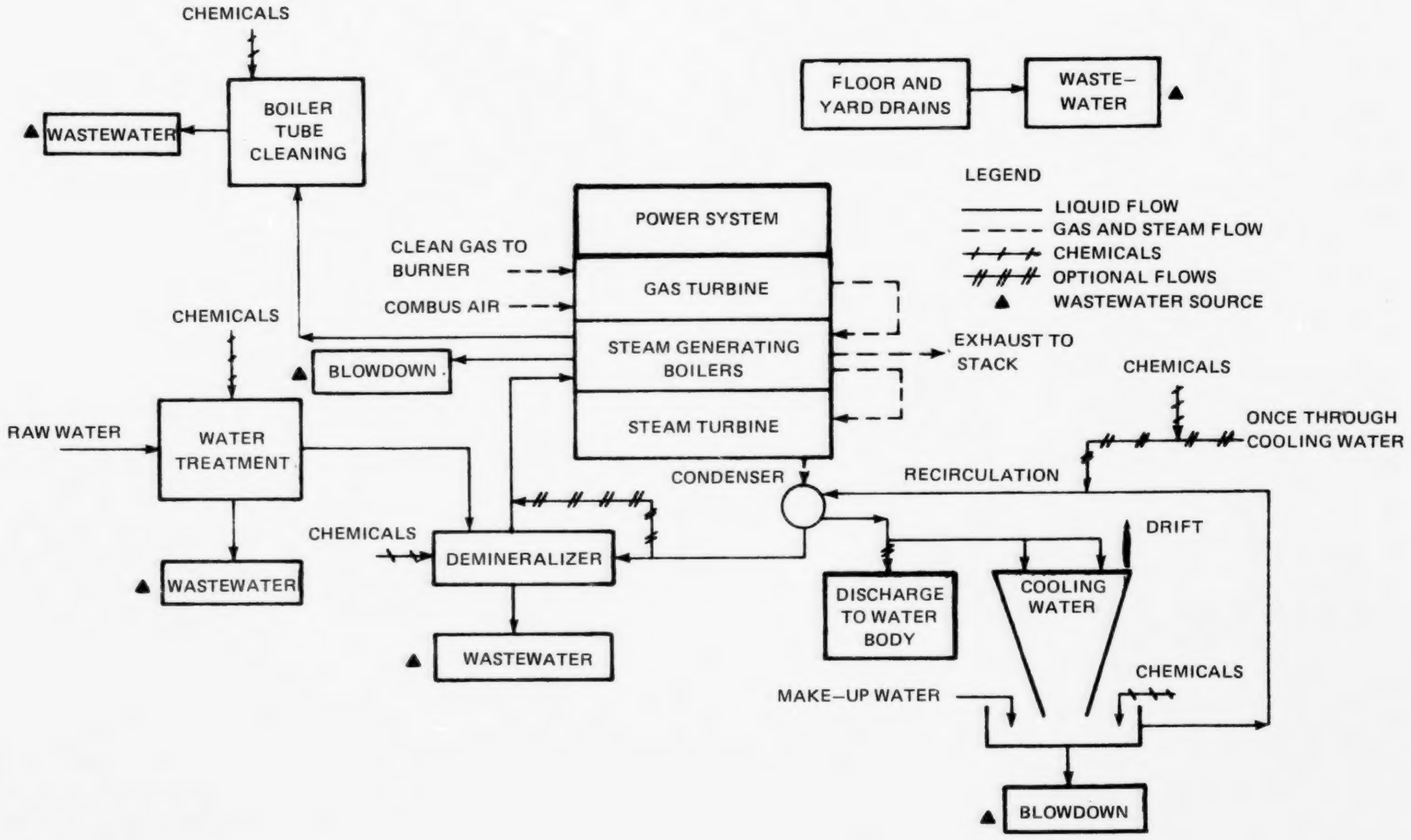
CHEMICALS
BOILER TUBE CLEANING
WASTEWATER
FLOOR AND YARD DRAINS
WASTE– WATER
LEGEND
LIQUID FLOW
GAS AND STEAM FLOW
CHEMICALS
OPTIONAL FLOWS
WASTEWATER SOURCE
POWER SYSTEM
CLEAN GAS TO BURNER
GAS TURBINE
COMBUS AIR
CHEMICALS
STEAM GENERATING BOILERS
EXHAUST TO STACK
BLOWDOWN
CHEMICALS
STEAM TURBINE
RAW WATER
WATER TREATMENT
ONCE THROUGH COOLING WATER
CONDENSER
RECIRCULATION
DRIFT
CHEMICALS
DEMINERALIZER
WASTEWATER
DISCHARGE TO WATER BODY
COOLING WATER
WASTEWATER
CHEMICALS
MAKE–UP WATER
BLOWDOWN

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FIG. 10 –5

SCHEMATIC REPRESENTATION OF WASTE WATER STREAMS AND ANTICIPATED TREATMENT

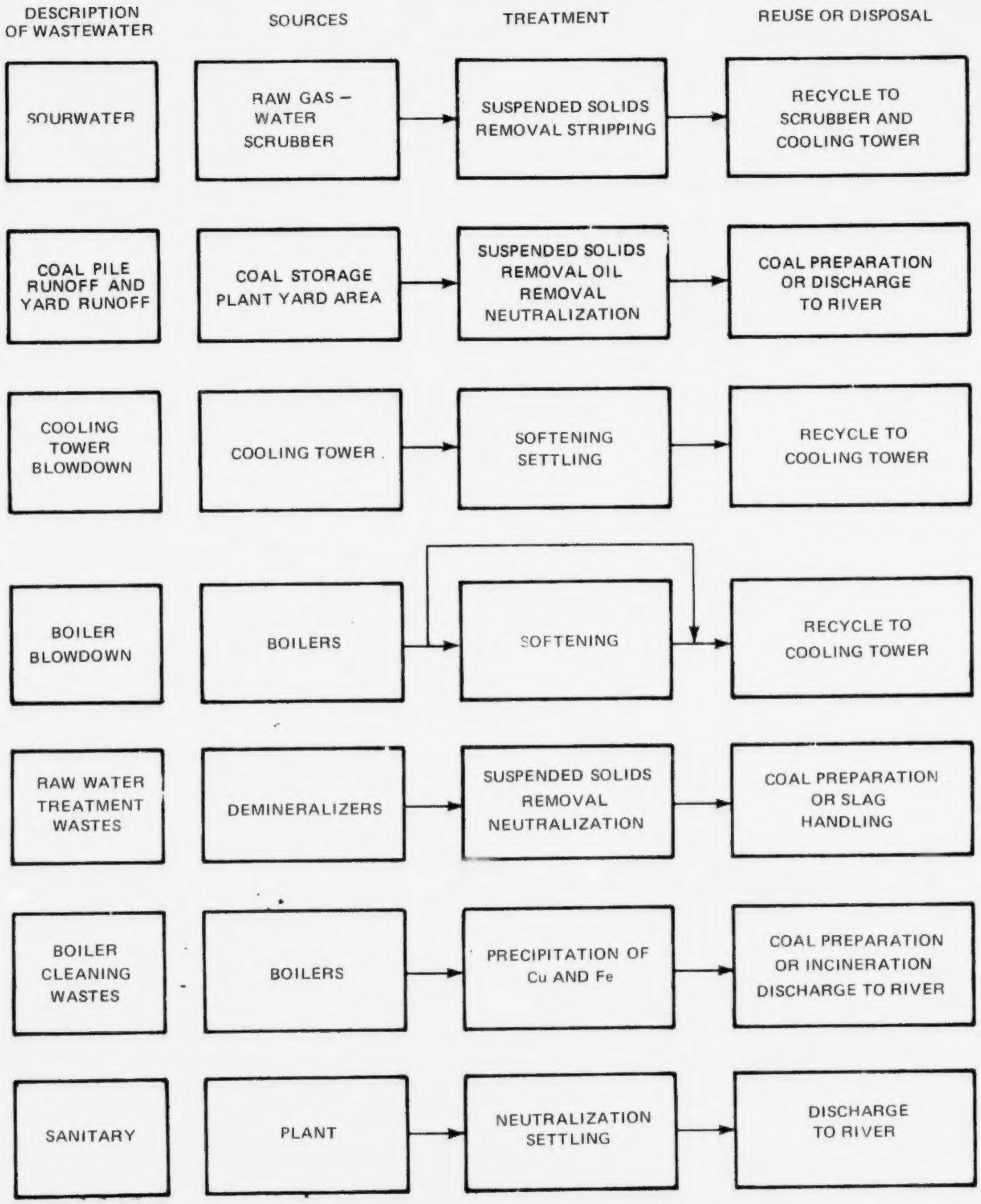
DESCRIPTION OF WASTEWATER
SOURCES
TREATMENT
REUSE OR DISPOSAL
SOURWATER
RAW GAS – WATER SCRUBBER
SUSPENDED SOLIDS REMOVAL STRIPPING
RECYCLE TO SCRUBBER AND COOLING TOWER
COAL PILE RUNOFF AND YARD RUNOFF
COAL STORAGE PLANT YARD AREA
SUSPENDED SOLIDS REMOVAL OIL REMOVAL NEUTRALIZATION
COAL PREPARATION OR DISCHARGE TO RIVER
COOLING TOWER BLOWDOWN
COOLING TOWER
SOFTENING SETTLING
RECYCLE TO COOLING TOWER
BOILER BLOWDOWN
BOILERS
SOFTENING
RECYCLE TO COOLING TOWER
RAW WATER TREATMENT WASTES
DEMINERALIZERS
SUSPENDED SOLIDS REMOVAL NEUTRALIZATION
COAL PREPARATION OR SLAG HANDLING
BOILER CLEANING WASTES
BOILERS
PRECIPITATION OF Cu AND Fe
COAL PREPARATION OR INCINERATION DISCHARGE TO RIVER
SANITARY
PLANT
NEUTRALIZATION SETTLING
DISCHARGE TO RIVER

TABLE 10-4

BY-PRODUCT WATER ANALYSIS FROM SYNTHANE GASIFICATION OF VARIOUS COALS

mg/l (except pH)

	Coke Plant	Illinois No. 6 Coal	Wyoming Subbituminous Coal	Illinois Char	North Dakota Lignite	Western Kentucky Coal	Pittsburgh Seam Coal
pH	9	8.6	8.7	7.9	9.2	8.9	9.3
Suspended Solids	50	600	140	24	64	55	23
Phenol	2,000	2,600	6,000	200	6,000	3,700	1,700
COD	7,000	15,000	43,000	1,700	38,000	19,000	19,000
Thiocyanate	1,000	152	23	21	22	200	188
Cyanide	100	0.6	0.2	0.1	0.1	0.5	0.6
NH_3	5,000	8,000 (1)	9,520	2,500	7,200	10,000	11,000
Chloride	-	500	-	31	-	-	-
Carbonate	-	6,000 (2)	-	-	-	-	-
Bicarbonate	-	11,000 (2)	-	-	-	-	-
Total sulfur	-	1,400 (3)	-	-	-	-	-

(1) 85 percent free NH_3

(2) Not from same analysis

(3) $S^{=}$ = 400

$SO_3^{=}$ = 300

$SO_4^{=}$ = 1,400

$S_2O_3^{=}$ = 1,000

TABLE 10-5

GASIFICATION WASTEWATER - ESTIMATED COMPOSITION

(pH = 8.4)(1)

Constituent	Concentration mg/l
Phenols	420
Other Organics	3,000
Thiocyanates	2
Hydrogen Sulfide	12
Ammonia	50
Hydrogen Cyanide	5
Total Dissolved Solids	930
Chloride	300
Calcium	2
Ferrous or Ferric Ions	1

(1) The gasification system is not specified

Cooling Tower Blowdown Characteristics--

In the operation of a closed cooling system, with wet cooling towers, the warm circulating water returning to the cooling system is cooled by evaporating a small fraction of it. The amount of water evaporated is a function of the temperature change of the water between the inlet and outlet of the cooling system; approximately one percent of the circulating water is evaporated for each 10F drop, assuming a latent heat for water of 1000 Btu/lb.

Additional water is lost to the atmosphere when water is entrained in the air draft (drift loss). Drift losses in modern mechanical-draft towers average 0.005 percent of the cooling-water circulating rate (0.002 percent in modern natural-draft towers), while in older and less efficient towers it can be as much as 0.2 percent of the circulating rate (Reference 10-4).

Because of the water losses from evaporation, the remaining water becomes more concentrated with dissolved solids. If the concentration level of any of the soluble salts exceeds its solubility level, the salt will precipitate. Some of the salts are characterized by reverse solubility, that is, their solubility decreases with increasing temperature. When cooling water saturated with such a salt is heated in the process condensers, the salt will deposit as a scale on the condenser tube walls and reduce heat transfer across the tubes.

Scale formation is controlled by discharging a blowdown stream from the cooling system to limit the concentration of dissolved solids. The amount of blowdown is a function of the number of concentration cycles, that is, the ratio of the content of the critical component in the circulating water to that in the makeup water. Since the drift loss is fixed for a given tower, the blowdown flow is varied to achieve the desired concentration cycles. Thus, for a tower with a 0.005 percent drift, blowdown may vary from 65 percent of makeup for a saltwater system to 20 percent of makeup for the typical freshwater tower.

A variety of chemical additives may be used to treat water circulating in the cooling system to control scaling, erosion, and fouling. These additives will appear in the blowdown along with matter originally present in the makeup stream. Biological growth in the circulating water is usually inhibited by chlorinating the water. Sulfuric acid is often added to cooling water to increase the solubility of the dissolved solids and to lower the make-up requirements due to blowdown. Pentachlorophosphate is sometimes added to inhibit fungi attack on wooden cooling towers.

There may be particular problems associated with leakage from the high-pressure gas process train into the cooling system. Such leakages, if they occur, will also be present in the cooling system blowdown.

Typical cooling tower blowdown characteristics, obtained from an operating power plant in Sioux Falls, South Dakota, are given in Table 10-6.

TABLE 10-6

COOLING TOWER BLOWDOWN CHARACTERISTICS

Characteristic[(1)]	Cooling Tower Blowdown
pH	8.7
Temperature (Winter), °F	69
Temperature (Summer), °F	84
Alkalinity (as $CaCO_3$)	123
BOD 5-Day	4.1
COD	2.7
Total Dissolved Solids	1,293
Total Suspended Solids	37
Ammonia (as N)	0.03

(1) Concentrations in ppm, except pH.

Boiler Blowdown Characteristics--

The formation of scale also is a major problem associated with the operation of boilers or waste heat recovery systems. The primary cause of scale formation is the reverse solubility of many of the scale forming salts. The higher the temperature and pressure of boiler operation, the more insoluble become the salts which form scale. Calcium and magnesium salts are the most common ingredients of boiler scales. Calcium deposition is primarily due to the thermal decomposition of calcium bicarbonate according to the following equation:

$$Ca(HCO_3)_2 \quad \rightarrow \quad CaCO_3 + CO_2 + H_2O \; . \qquad (10\text{-}2)$$

Deposits of iron oxide, copper oxide, and other metallic oxides are frequently found in boilers operating with feedwater having high dissolved oxygen content. These deposits are left by corrosion resulting from dissolved oxygen and carbon dioxide.

Boiler blowdown is the most widely used control method for scale formation. The amount of blowdown required is a function of the allowable concentration of scale forming or other undesirable components in the boiler and the degree to which the make-up water is purified. High-pressure boilers have quite stringent contaminant limits. For example, the allowable concentration of silica varies from 125 ppm at pressures under 300 psi down to 0.5 ppm at pressures in excess of 2000 psi. As a result, the allowable number of concentrations can be quite low in a high pressure steam system. At pressures above 600 psi, silica present in the boiler will vaporize along with other contaminants and escape with the steam. To eliminate silica condensation and resultant fouling of the turbine, it is necessary to maintain extremely low silica concentrations in the boiler which can result in a high amount of blowdown. Other methods, such as steam washing, can be used to reduce the contaminant vapor content of the steam permitting higher boiler water concentrations and reducing the required blowdown quantity or makeup water quality. In the steam washing method, high-pressure steam bypasses the turbine for a short period.

Boiler blowdown contains all of the boiler feedwater additives as well as the soluble matter originally present in the boiler feedwater. Scale formation is usually inhibited by adding chemicals, such as phosphates, which precipitate scale-forming salts to form sludge. Chelating agents are also widely used. They form complex compounds with scale-forming metal ions, thus increasing their solubility. Sodium sulfite or hydrazine is often added to boiler feedwater to inhibit corrosion from dissolved oxygen.

Boiler blowdown is alkaline with a pH of 9.5 to 10 for hydrazine treated water and a pH of 10 to 11 for phosphate treated water. Hydrazine treated boilers produce blowdown containing up to 2 ppm ammonia, those treated with phosphate may contain up to 50 mg/ℓ phosphate and up to 100 mg/ℓ hydroxide alkalinity.

Water Treatment Wastes--

Water treatment waste streams are described by pH, suspended solids concentration, and concentration parameters typical of the processes involved or toxic elements involved in the processes.

Clarification wastes consist of clarifier sludge and filter washes. Clarifier sludge could be either alum or other ionic based sludges from coagulant chemicals. If the clarifier is used for both, lime softening and clarification, the sludge will contain calcium carbonate and magnesium hydroxide. Filter washes contain suspended solids occurring either as light carryover floc from the clarifier or occurring naturally in unclarified raw water. The underflow from clarifiers used in treating makeup water contains from 0.5 percent to 5 percent solids. This underflow will go to a sludge thickener, where the solids content increases to 10-15 percent. The supernatant water from the thickener is returned to the clarifier inlet.

Ion exchanger wastes may be either acidic, alkaline or neutral. Usually, such wastes do not contain suspended matter. They may, however, contain calcium sulfate and calcium carbonate precipitates because of the common ion effect.

Equipment Cleaning Wastes--

A variety of cleaning formulations are used to clean scale and corrosion deposits from boilers and condensers. The cleaning procedure is usually dependent on the composition of the surface-adhering materials. Cleaning solutions are usually grouped by composition into three principal categories. The first category includes the alkaline cleaning mixtures with an oxidizing agent and a copper chelating compound, usually ammonia, for copper removal. The oxidizing compound converts metallic copper deposits to a divalent copper ion which then reacts with ammonia to form a soluble complex. The wastewater effluents contain an ammonium ion, oxidizing agents, high levels of dissolved copper and iron, and have high alkalinity.

The second category includes acidic cleaning mixtures. These mixtures effectively remove scale caused by water hardness. They contain a strong acid and a fluoride salt to remove silica. Waste streams from such mixtures are usually acidic and may contain phosphates, fluorides and BOD, as well as large quantities of iron, copper, and hardness forming salts.

The last group of formulations includes solutions containing alkaline chelating agents and anticorrosion additives. These cleaning mixtures may be used alone or after acid cleaning to neutralize residual acidity as well as to remove additional amounts of scale-forming materials. Their use generates wastewater containing alkalinity, BOD, phosphate, and scale-forming components.

In addition to the three categories of cleaning formulas above, several proprietary formulations have been developed and are manufactured by companies specializing in cleaning chemicals. Most of these chemicals are similar to those described; the resulting wastes contain alkalinity, BOD, phosphate, ammonium compounds, and scale-forming compounds such as iron, copper, and hardness-forming salts.

Floor and Yard Drains--

The flood drains generally contain dust and fines and floor scrubbing detergents. This stream also contains lubricating oil or other oils washed away during equipment cleaning, oil leakage from pump seals, and oil collected from spillage around the storage tank area of oil processing gasifiers.

Solid Wastes

Power generation by any of the seven integrated power systems considered in this study will result in appreciable quantities of various solid wastes. This discussion is confined to wastes exiting the coal preparation, gasification, gas cleanup, power system and auxiliary facilities. Therefore, construction debris, solids from sewage systems, garbage disposals, and solid waste problems that are part of developmental efforts, are not addressed in this report. Solids formed as a result of control technology selection such as water treatment sludges, elemental sulfur and recovered entrained particulates are discussed later in this section under the heading of Residuals.

Figure 10-6 shows the points at which solids exit a generalized integrated system. Mine tailings such as pyrite and other mineral matter will be removed during coal processing. Intermittent losses of coal solids will occur during preparation for the gasifier. Solid wastes exiting the coal preparation unit were discussed earlier in this section. The mineral residues of gasified coal will exit the gasifiers as slag. Small quantities of dissolved gases may be present in the slag. Spent acceptors, such as dolomite in the Conoco gas cleanup process, will exit the gas cleanup operations.

Tables 10-7 and 10-8 show the output rates for some solid wastes generated in the specified integrated systems.

Based on current data it is impossible to give a detailed description of the gasifier slags. The variety of possible ash characteristics is as unlimited as the number of possible coal compositions. Furthermore, gasifier operating conditions will influence the slagging characteristics of the ash contained in any specified coal feed.

Table 10-9 cites three ash characterizations. The first two are general descriptions of coal ash formed by coal combustion. The third column of Table 10-9 characterizes the ash exiting a Lurgi gasifier. The distribution of mineral matter is similar to that of the general description although some constituents are present in quantities that do not fall in the ranges given for the general ash characteristics.

Solid Waste Disposal--

With the possible exception of spent dolomite, the solid waste generated during integrated power generation will require no additional treatment prior to ultimate disposal. Solids from air and water control technologies, however, may require further treatment prior to disposal.

FIG. 10–6

SOLID WASTES EXITING A GENERALIZED INTEGRATED PLANT

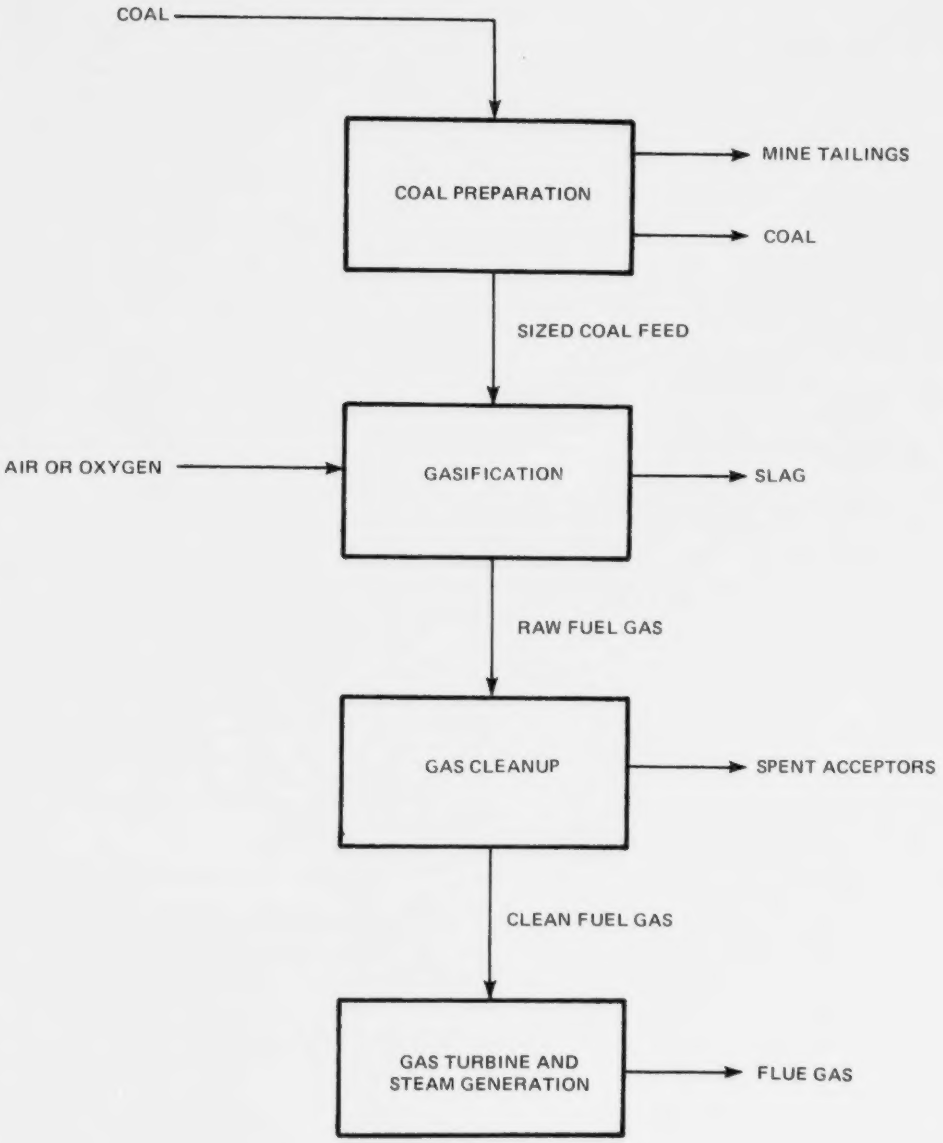
COAL
COAL PREPARATION
MINE TAILINGS
COAL
SIZED COAL FEED
AIR OR OXYGEN
GASIFICATION
SLAG
RAW FUEL GAS
GAS CLEANUP
SPENT ACCEPTORS
CLEAN FUEL GAS
GAS TURBINE AND STEAM GENERATION
FLUE GAS

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TABLE 10-7

SUMMARY OF SOLIDS EXITING THE BCR AND IGT COMBINED CYCLES

Configuration/Type of Solid Waste	Quantity of Waste, lb/hr
BCR/Selexol/Air and Oxygen	
Slag	70,590
IGT/Selexol/Air	
Slag	73,640
BCR/Conoco/Air	
Slag	70,590
Dolomite Slurry (Total)	62,200
Water	34,170
Dolomite	25,660
Inerts	2,370
Total Solid Wastes	132,800*
BCR/Conoco/Oxygen	
Slag	70,590
Dolomite Slurry (Total)	57,360
Water	31,510
Dolomite	23,640
Inerts	2,180
Total Solid Wastes	127,950
IGT/Conoco/Air	
Slag	73,640
Dolomite (Total)	61,520
Water	33,800
Dolomite	25,380
Inerts	2,340
Total Solid Wastes	136,700

*Totals are not exact due to rounding of all values to four significant digits.

TABLE 10-8

DESCRIPTION OF SLAG AND SALT SLURRY EXITING THE MOLTEN SALT GASIFIER

Slurry Constituents	Quantity, lb/hr
Aqueous Phase	
Na_2CO_3 (aq)	4,840
$NaHCO_3$ (aq)	9,873
NaHS (aq)	2,150
H_2O (l)	81,386
Total aqueous phase	98,249
Na_2CO_3 (s)	20,235
$NaHCO_3$ (s)	5,361
Sand (s)	64
Slag (s)	72,589
Total solid phase	98,248
Total solid waste	196,497

TABLE 10-9

CONSTITUENTS OF COAL ASH

Constituent	Percent (1)	Percent (1)	Percent (2)
SiO_2	30-50	20-60	59.0
Al_2O	20-30	10-35	23.7
Fe_2O_3	10-30	5-35	4.7
TiO_2	0.4-1.3	0.5-2.5	0.9
CaO	1.5-4.7	1-20	3.7
MgO	0.5-1.1	0.3-4	0.9
Na_2O	0.4-1.5	1-4	1.4
K_2O	1.0-3.0	1-4	0.8
SO_3	0.2-3.2	0.1-12	NS
C and volatiles	0.1-4.0	NS	5.0
P	0.1-0.3	NS	NS
B	0.1-0.6	NS	NS
U and Th	0.0-0.1	NS	NS
Cu	trace	NS	NS
Mn	trace	NS	NS
Ni	trace	NS	NS
Pb	trace	NS	NS
Zn	trace	NS	NS
Sr	trace	NS	NS
Ba	trace	NS	NS
Zr	trace	NS	NS

NOTE: NS denotes value not specified.

(1) Conventional coal-fired boiler

(2) Lurgi gasifier

Landfilling is the primary technique used for ultimate disposal of solid wastes. Ash disposal via ponds was considered, but is not discussed here because this method of ultimate solid waste disposal is highly site specific. Unless large amounts of land are available adjacent to the other facilities, the costs associated with waste transport become prohibitive. Landfilling is in common use for both industrial and municipal solid waste disposal. If properly designed, landfills are also suitable for disposal of hazardous materials. Landfill designs are site specific; local soil, rainfall, and water table characteristics are among the major design parameters. The quantity and physical and chemical properties of the solids to be landfilled are important process specific variables.

Despite the fact that every landfill design is unique, certain generalizations can be made. Generally, to be economical, landfills should be located within ten miles of the point of waste generation. Landfills should not interface local water tables. A minimum of not more than two feet of soil between the landfill and anticipated high groundwater level is required in several states. For this reason, landfill capacities should be determined by usable land volume rather than land area.

A regular program of local air and water monitoring around the landfill should be established. Materials from the combined cycle power systems deposited in a landfill are poorly characterized. Thus, the precise effects of these wastes are unknown. As a result, a regular monitoring program and detailed records of all landfilling and monitoring activities are essential.

One alternative to landfilling of solid wastes is to develop ways to use them beneficially as raw materials. Slag could be used a constituent of construction materials such as concrete building blocks and asphalt. Another potential application is to use slag as an abrasive. Fragmented slags from other industries have been used in sand blasters to remove paint from marine vessels.

RESIDUALS

Residuals are generated as a result of applications of control and disposal technologies. The following paragraphs briefly discuss possible adverse environmental effects which could be caused by these wastes. Where necessary, additional control and disposal practices are identified.

Gaseous, liquid, and solid residuals exit the systems controlling emissions to air. Particulates removed from cyclones can be returned to the gasifier or landfilled with solids from the slag removal system. Two residual streams exit the Claus process. Spent Claus catalyst may require chemical stabilization to neutralize ammonia, organic sulfur compounds, and carbonaceous compounds prior to disposal in the landfill to avoid leaching problems. Claus sulfur may be sold as a by-product or landfilled. New uses for sulfur are being developed in order to increase demand for this commodity. Potential applications include sulfur-based fertilizers, construction materials and insulating foams.

Additional sulfur, of high purity, is recovered by the Beavon process and can be marketed or disposed of with Claus sulfur. The Beavon oxidizer vent is a clean gas stream, containing nitrogen, oxygen and water. Minor quantities of ammonia may be present in the Beavon feed gas. If present, it will exit with the oxidizer vent gas. Two water streams from the Beavon unit, the process condensate and the Stretford sorbent blowdown, will require further treatment. Stripping out H_2S from the condensate stream should make that stream suitable for reuse in the plant. The Stretford sorbent blowdown stream should be sent to the wastewater treatment system to remove thiosulfates, thiocyanates and sulfates.

Black water, a wastewater generated by the liquid-phase Claus process, is found in the Conoco configurations. This water can be used to quench gasifier slag and to remove spent dolomite in slurry form.

The water and wastewater treatment facilities will generate solid residuals. Solid wastes from water and wastewater treatment processes are collectively referred to as sludges. There are two types of sludge in water and wastewater. The first is suspended solid materials which form sludges through the mechanisms of coagulation and sedimentation. The second is dissolved chemical salts. Precipitation of these salts forms sludges.

The physical and chemical properties of a sludge depend upon the characteristics of the feed water and the treatment operation selected. Clarifier sludges from raw water treatment are either alum or iron salt sludges, depending on the type of chemical coagulant selected. For example, alum sludges consist of aluminum hydroxide, inorganics such as clay and sand particles, and organic matter including plankton. The sludge from a lime soda softening clarrifier is calcium carbonate, and also contains magnesium hydroxide, iron or aluminum hydroxide, suspended mineral matter, and organics present in lesser amounts. Treatment of coal pile runoff and floor and yard drainage also produces sludge. The primary chemical salts present in such a sludge are calcium carbonate and hydroxides of iron, aluminum, chromium, zinc and manganese. Some coal fines, dust, and oil also may be present. Water demineralization also may produce sludges, calcium sulfate being the primary constituent. These sludges can be disposed of by landfilling. Mixing the sludge from the softening unit with slag should further reduce the risk of leaching of undesirable materials to local groundwaters.

TRACE ELEMENTS

The distribution of trace elements in Illinois No. 6 coal, and the forms in which they possibly occur in coal were described earlier in this section. Trace elements are present in less than 1 percent concentrations. However, they must be considered from the standpoint of their potential environmental impact.

Even in trace amounts, very toxic substances could have adverse effects. Because of the large number of elements present, it is important to identify those most likely to be of environmental concern. Having done this, discussion of the environmental impact can focus on them.

Trace Element Evaluation

If trace elements or compounds of them are discharged to the environment as a result of integrated power system operations, they may cause adverse health or ecological effects. To aid in evaluating and ranking the elements or their compounds which may have adverse environmental impact, a convenient tool, suggested by the EPA Technical Project Officer, uses the methodology of Multimedia Environmental Goals (MEG's) (Reference 10-7). MEG's characterize environmental pollutants by postulating emission level goals and ambient level goals for air, water, and land. These goals are the result of consistent extrapolations which consider the following data: existing standards and criteria; threshold limit values; acceptable risk levels for human exposure to suspected carcinogens or teratogens; contamination considered reasonable for protection of ecosystems; and cumulative potential in living organisms.

Both acute and chronic effects are of interest for assessing the environmental impact of trace elements from integrated power systems. MEG's address acute and chronic effects in several ways including the following: Minimum Acute Toxicity Effluents (MATE's) are emission level goals which give estimates for concentrations of pollutants in undiluted emission streams that will not adversely affect those persons (for health effects) or ecological systems (for ecological effects) exposed for short periods of time; estimated Permissible Concentrations (EPC's) are ambient level goals based on standards or criteria, toxicity and carcinogenic or teratogenic potential. Toxicity-based EPC's represent concentrations of toxic substances in emission streams which, after dispersion, will not cause the level of contamination in the ambient media to exceed a safe continuous concentration for exposure of persons (for health effects) or ecological systems (for ecological effects).

In the absence of data from actual gasification plants it has been assumed that all the trace elements present in coal are discharged to air, water and land in the same proportions as they are present in coal. According to the method suggested by the Project Officer, dividing the concentration of trace elements in coal by the corresponding concentration for these elements indicative of adverse health or ecological effects will help in establishing rank values for the elements. Values of the two MEG's described above, the MATE's and EPC's for adverse health effects through air and water have been used. The same could be done with ecological MATE's and EPC's.

Water EPC's are confused by the many standards set for water. In cases where a standard exists, the EPC is the standard and the MATE is exactly 5 times the EPC. In other cases, where no standard exists, EPC and MATE values are not simply related. In present lists of MEG values, water and land values are proportional. Thus, rankings of a given trace element distribution are identical for land and water and are designated by water-land. The MATE and EPC values for the trace elements in air and water are shown in Table 10-10. The method employed for ranking the trace elements in terms of health effects (or ecological effects, if desired) is a four step procedure as follows:

TABLE 10-10

MINIMUM ACUTE TOXICITY EFFLUENT AND ESTIMATED PERMISSIBLE CONCENTRATION VALUES FOR AIR AND WATER

Element	Coal Conc. ppm	MATE Air ($\mu g/m^3$)	MATE Water ($\mu g/l$)	EPC** Air ($\mu g/m^3$)	EPC** Water ($\mu g/l$)
Ag	0.11	10	250	0.024	50*
As	24.0	2	250	0.005	50*
B	200.0	3,100	47,000	7.4	43
Ba	31.0	500	5,000	1.0	1,000*
Be	1.0	2	30	0.01*	4
Cd	0.89	10	50	0.12	10*
Co	3.6	50	750	0.12	0.7
Cr	15.0	1	250	0.12	50*
Cu	19.0	200	5,000	0.5	1,000*
Ge	4.3	560	3,400	1.3	8
Hg	0.12	50	10	0.1	2*
Mn	48.0	5,000	250	12.0	50*
Mo	7.0	5,000	75,000	12.0	70
Ni	15.0	15	230	0.24	1.4
Pb	11.0	150	250	0.36	50*
Sb	1.0	500	7,500	1.2	7
Se	13.0	200	50	0.5	10*
Sr	37.0	3,000	4,000	5.5	27
V	17.0	500	2,500	1.2	7
Zn	49.0	4,000	25,000	9.5	5,000*
Mg	570.0	8,000	90,000	14.0	83
Ti	700.0	6,000	90,000	14.0	83

---

*Standard

**Toxicity Based

Step 1. Obtain the ratio values of coal concentration and MATE, and coal concentration and EPC for all the trace elements of interest as revealed by analyses of the coal and corresponding MEG values. In terms of simple equations Step 1 can be stated as:

$$(RM)i = \frac{(TEC)i}{(MATE)i} \tag{10-3}$$

$$(RE)i = \frac{(TEC)i}{(EPC)i} \tag{10-4}$$

Where:
(RM)i: Ratio value for trace element i using MATE
(RE)i: Ratio value for trace element i using EPC
(TEC)i: Concentration of trace element i in coal

Step 2. Give ranking numbers (MRANK)i and (ERANK)i to the elements on the (RM)i and (RE)i lists, respectively, based on the values, with rank one given to the smallest ratio value. Thereby (MRANK)i and (ERANK)i will be obtained for all the trace elements. (MRANK)i denotes the rank of trace element i when MATE is used and (ERANK)i denotes the rank of trace element i when EPC value is used.

Step 3. Add the (MRANK)i and (ERANK)i to obtain (TVAL)i for all trace elements. In effect, this weights equally the measures of acute and chronic adverse health (or ecological) effects to obtain a single ranking:

$$(TVAL)i = (MRANK)i + (ERANK)i. \tag{10-5}$$

Step 4. Give a new rank (NRANK)i based on the (TVAL)i values, with one given to the element having highest value of TVAL. This corresponds to the highest overall ranking for adverse health (or ecological) effects.

Employing the four steps above on the coal considered in this study and using MATE and EPC values such as in Table 10-10, (RM)i and (RE)i values for air and water are calculated. Some of these are shown in Table 10-11. (MRANK)i or (ERANK)i values are shown in the left column. The elements are displayed in descending order of the rank which resulted. Steps 3 and 4 are then done to obtain (TVAL)i, and based on these values for the elements, the rank (NRANK)i is given to each element. The (TVAL)i and (NRANK)i values for air and water are shown in Table 10-12. From this table it is evident that arsenic, nickel, and titanium rank in the top five for both air and water-land. Titanium's rank is due mainly to its high concentration in coal. It should present no problem if it remains in the slag. Chromium is second on the list for air. However, since chromium, a nonvolatile element, should also remain in the slag, it may not escape, either with any air emission that may occur or with the sour water generated during scrubbing. Nonetheless, chromium, if leached from the slag, will affect the ground water quality. Elements present in the slag may be in the form of oxides, sulfides, or carbonates; these compounds tend to generate leachate of the elements. Cadmium and lead are ranked

TABLE 10-11

ENVIRONMENTAL METHODOLOGY FOR RANKING TRACE ELEMENTS IN AIR AND WATER

(MRANK)i or (ERANK)i	Air				Water			
	(RM)i		(RE)i		(RM)i		(RE)i	
22	Cr	15.0	As	4,800.0	Se	0.26	Ni	10.71
21	As	12.0	Cr	125.0	Mn	0.192	Ti	8.43
20	Ni	1.0	Be	100.0	As	0.096	Mg	6.87
19	Be	0.5	Ni	62.5	Ni	0.0652	Co	5.14
18	Ti	0.12	Ti	50.0	Cr	0.06	B	4.65
17	Mg	0.095	Mg	40.7	Pb	0.044	V	2.43
16	Cu	0.095	Cu	38.0	Be	0.0333	Sr	1.37
15	Cd	0.089	Ba	31.0	Cd	0.0178	Se	1.3
14	Pb	0.074	Pb	30.56	Hg	0.012	Mn	0.96
13	Co	0.072	Co	30.0	Ti	0.0078	Ge	0.54
12	Se	0.065	B	27.03	V	0.0069	As	0.48
11	B	0.0645	Se	26.0	Mg	0.0063	Cr	0.3
10	Ba	0.062	V	14.17	Ba	0.0062	Be	0.25
9	V	0.034	Cd	7.42	Co	0.0048	Pb	0.22
8	Sr	0.0123	Sr	6.73	B	0.0043	Sb	0.14
7	Zn	0.0122	Zn	5.16	Cu	0.0038	Mo	0.1
6	Ag	0.011	Ag	4.58	Zn	0.0019	Cd	0.09
5	Mn	0.0096	Mn	4.0	Ge	0.0013	Hg	0.06
4	Ge	0.008	Ge	3.31	Sr	0.008	Zn	0.01
3	Hg	0.0024	Hg	1.2	Ag	0.004	Ba	0.003
2	Sb	0.002	Sb	0.83	Sb	0.0001	Ag	0.0022
1	Mo	0.0014	Mo	0.58	Mo	0.0009	Cu	0.0019

TABLE 10-12

ENVIRONMENTAL RANKING OF TRACE ELEMENTS IN AIR AND WATER-LAND

Air			Water-Land	
(TVAL)i	Element	(NRANK)i	Element	(TVAL)i
43	As	1	Ni	41
42	Cr	2	Se	37
40	Be	3	Mn	35
39	Ni	4	Ti	34
36	Ti	5	As	32
34	Mg	6	Mg	31
32	Cu	7	V	29
28	Pb	8	Cr	29
26	Co	9	Co	28
25	Ba	10	Pb	26
24	Cd	11	Be	26
23	Se	12	B	26
23	B	13	Cd	21
19	V	14	Sr	20
16	Sr	15	Hg	19
14	Zn	16	Ge	18
12	Ag	17	Ba	13
10	Mn	18	Sb	10
8	Ge	19	Zn	10
6	Hg	20	Cu	8
4	Sb	21	Mo	8
2	Mo	22	Ag	5

in the middle for both air and water-land. Mercury, near the bottom of the list, is volatile and will likely appear in quench water. It possibly could gain a higher ranking if a volatility factor is considered for water-land ranking. Vanadium shows higher environmental rank through water-land than through air. Boron appears in the middle of the list for both air and water.

These rankings are useful for a first cut at relative environmental effects caused by various elements. The method gives a preliminary comparison among different elements which includes both their concentration in coal and their MEG values. The ranking could be misleading if all considerations are not taken into account, such as with the case of chromium.

Considering the volatilities of trace elements is a valuable refinement to the ranking procedure. The (RM)i and (RE)i values of trace elements designated in Table 10-11 can be multiplied by the corresponding volatility factors. The percentage change values for Illinois No. 6 coal in Table 10-13 were taken to be representative values for volatilities. These are taken from Reference 10-3 and are a summary of analytical results of feed and residue samples of coal hydrogasified in a bench-scale unit. The same procedure as outlined previously is then followed to establish the ranks of trace elements in air and water-land. These modified ranks are shown in Table 10-14. As expected, mercury ranks significantly higher for water-land. Arsenic, cadmium, lead, beryllium mercury, nickel, vanadium, and boron appear near the top of either air rankings or water-land rankings. Chromium has a low effect through air or water-land and ranks near the bottom of the lists. As stated earlier, this reflects the nonvolatile nature of chromium which causes it to remain in slag. This modified ranking further justifies the focus on the trace elements which were selected and asterisked in Table 10-14. When volatility is considered, it appears that selenium would be a logical choice for addition to the list of those to be more critically examined.

Environmental Impact of Trace Elements

The gasifier, an enclosed vessel, discharges no pollutants directly to the environment. However, the fate of the trace elements, whether they leave in the gasifier raw gas or in the gasifier ash/slag is controlled by factors such as reactor configuration and bed type, extent of pretreatment, type of coal feed system, and operating conditions. The trace elements that leave the gasifier in the raw fuel gas and hot ash/slag streams may find their way into the atmosphere and groundwater through various vent gas, water, and solid discharges from the downstream processing steps including the water treatment reuse system, the sulfur recovery unit, and the slag quench.

The possible discharge sources of trace elements are shown in Figure 10-7. Particulates discharged to air from coal storage and coal preparation will probably have the same trace element distribution as the parent coal. Water from the same sources will contain suspended coal solids; again, the trace element distribution in these solids will be the same as in the parent coal. The refuse from coal preparation will be mainly coal associated mineral matter, with a few trace elements associated with the mineral matter.

TABLE 10-13

TRACE ELEMENT DISTRIBUTION ILLINOIS NO. 6 COAL

Element	Feed, ppm(1)	% Change(2)	Final Residue, ppm(3)
Antimony	1.0	-28	0.72
Arsenic	24	-33	16
Barium	31	0	31
Beryllium	1.0	-24	0.76
Bismuth	1.1	-51	0.54
Boron	200	-10	180
Cadmium	0.89	-76	0.21
Calcium	3500	-34	2,300
Chlorine	2,300	-74	590
Chromium	15	0	15
Cobalt	3.6	0	3.6
Copper	19	0	19
Fluorine	61	-26	45
Germanium	4.3	-9	3.9
Iron	14,000	-8	13,000
Lead	11	-47	5.8
Lithium	33	0	33
Magnesium	570	+2	580
Manganese	48	-19	39
Mercury	0.12	-96	0.005
Molybdenum	7.0	-3	6.8
Nickel	15	-7	14
Nitrogen	10,000	-76	2,400
Potassium	1,700	0	1,700
Samarium	0.74	0	0.74
Selenium	13	-42	7.5
Silicon	20,000	0	20,000
Silver	0.11	-67	0.036
Sodium	1,400	+7	1,500
Strontium	37	0	37
Sulfur	38,000	-79	7,800
Tellurium	8.1	-41	4.8
Thorium	1	--	1
Tin	2.0	-45	1.1
Titanium	700	-3	679
Vanadium	17	-18	14
Ytterbium	0.56	-9	0.51
Zinc	49	-27	36
Zirconium	35	0	35

NOTES:

1. Feed composition is the same as in Table 10-1.
2. Average values of analytical results from two similar gasification runs in Hygas.
3. Minus sign indicates loss and plus sign indicates gain of any given element. The percent losses shown for PDU (Hygas) runs are good to ± 6 percent (i.e., losses less than 10-12% are not significant).

TABLE 10-14

MODIFIED ENVIRONMENTAL RANKING OF TRACE ELEMENTS IN AIR AND WATER-LAND

Air		Water-Land
Element	(NRANK)i	Element
*As	1	Se
*Be	2	Mn
*Pb	3	*As
*Ni	4	*Pb
*Cd	5	*Cd
Se	6	*B
Ag	7	*Hg
*B	8	*V
*V	9	Ti
Ti	10	*Be
Zn	11	*Ni
*Hg	12	Zn
Mn	13	Sb
Sb	14	Ag
Mo	15	Mo
*Cr	16	Mg
Mg	17	Co
Cu	18	*Cr
Co	19	Sr
Ba	20	Cu
Sr	21	Ba
Ge	22	Ge

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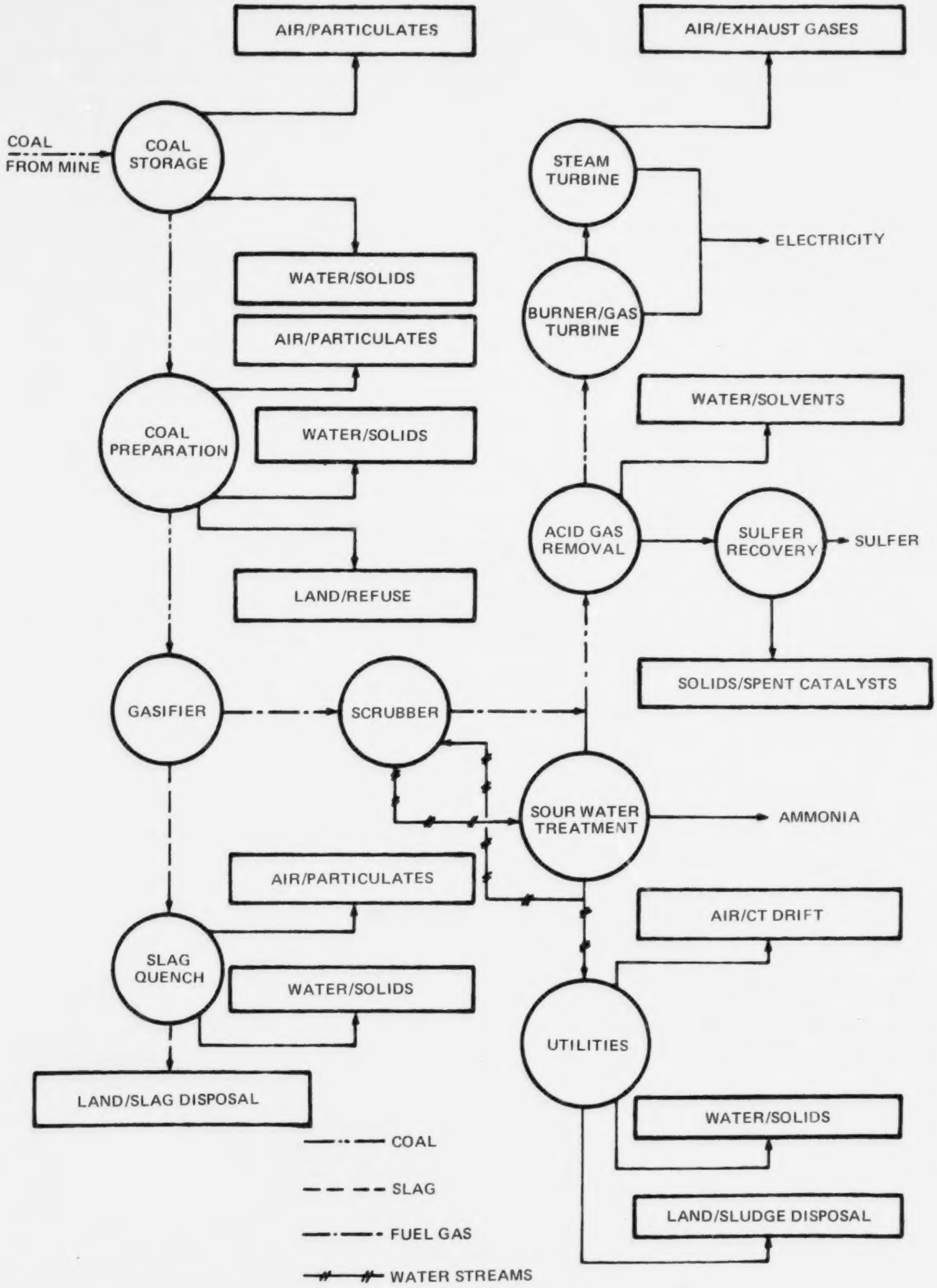
*Elements selected for focus (See Table 1-6).

FIG. 10 –7

POSSIBLE DISTRIBUTION OF TRACE ELEMENTS


AIR/PARTICULATES
AIR/EXHAUST GASES
COAL FROM MINE
COAL STORAGE
STEAM TURBINE
ELECTRICITY
WATER/SOLIDS
AIR/PARTICULATES
BURNER/GAS TURBINE
COAL PREPARATION
WATER/SOLIDS
WATER/SOLVENTS
ACID GAS REMOVAL
SULFER RECOVERY
SULFER
LAND/REFUSE
SOLIDS/SPENT CATALYSTS
GASIFIER
SCRUBBER
SOUR WATER TREATMENT
AMMONIA
AIR/PARTICULATES
AIR/CT DRIFT
SLAG QUENCH
WATER/SOLIDS
UTILITIES
LAND/SLAG DISPOSAL
WATER/SOLIDS
LAND/SLUDGE DISPOSAL
COAL
SLAG
FUEL GAS
WATER STREAMS

78–01–92–4

The coal conversion reaction takes place in a reducing atmosphere and may form compounds of the trace elements such as hydrides, carbonyls or sulfides which may be more volatile than the element. Some trace elements in the coal react with the organic matter in ash to form organometallic compounds. The high operating pressures in the gasifier may cause carbon monoxide to form carbonyls with iron and nickle. The formation of three typical compounds and their properties are shown in Table 10-15 (Reference 10-8). Many of the trace elements volatilize to varying extents during the high-temperature processing. The forms in which these elements appear in the raw gas are determined by their chemical form in the feed coal and by processing conditions. The predicted volatility of some elements is given in Table 10-16 from Reference 10-9. Elements having a high volatility and high toxicity can adversely affect the environment if released to the atmosphere. Such elements include mercury, selenium, arsenic, and lead. In the case of zinc, boron, and fluorine, the degree to which they volatilize has not yet been determined, but may be rather significant. If only 10 percent of these elements are volatilized, large quantities will be present in the raw gas from the gasifier since they are present in the coal in relatively higher concentratins than other trace elements.

It should be noted that volatilities estimated by the various researchers are based mainly on work done for the EPA at IGT (References 10-3 and 10-10). This work centered on the analysis of feed and residues from the various stages of the HYGAS process. It includes data from both bench scale and pilot plant work. Because gasifier conditions may vary widely from those of the HYGAS process, the form in which the trace elements appear will vary as will their associated volatilities. Nevertheless, these extrapolations are of interest and should be helpful in obtaining a proper perspective on the trace element problem.

Trace elements which are nonvolatile or are only partially volatile will leave the gasifier with the slag. Theoretically, if the trace element composition of the entering coal and the exiting slag is known, the trace elements leaving the gasifier with the raw gas can be estimated. However, trace element analysis is a difficult laboratory procedure; thus, the estimated raw gas composition from the analysis of Lurgi ash and feed coal does not correlate with the analytical trace element composition of the raw gas, as shown in Table 10-17 from Reference 10-11.

The slag from the gasifier is quenched and some of the trace elements in the mineral matter may dissolve in the quench water. The slag form of the mineral matter reduces the level of water contamination. A very low discharge of particulates to atmosphere from slag handling will occur. Slag is generally transported to the disposal site; further distribution of trace elements from slag will depend on the type of disposal method used. The trace elements in slag material are not likely to respond independently to weathering, leaching, burying or other chemical or physical processes; their behavior more likely will be tied directly to the behavior of the major mineral components.

TABLE 10-15

FORMATION OF TRACE ELEMENT COMPOUNDS

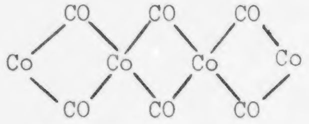
Compound, Structure	Synthesis Path	M.P., ^{o}C	B.P., ^{o}C	Comments
$Ni(CO)_4$ Nickel Carbonyl	$NiCl_2 \xrightarrow{NH_3CO} Ni(CO)_4$	-25	43	oxidizes slowly in air
$Fe(CO)_5$ Iron Carbonyl	$Fe + 5CO \xrightarrow[250^{o}C]{S} Fe(CO)_5$	-21	103	oxidizes readily in air
 Octa(carbonyl)-dicobalt	$CoCO_3 \xrightarrow[150^{o}C]{H_2,\ CO} Co_2(CO)_8$	51	52	unstable in air, has catalytic effects.

TABLE 10-16

PREDICTED VOLATILITIES OF TRACE ELEMENTS

Element	Possible % of Content of Coal Volatile [(1)]
Cl	90+
Hg	90+
Se	74
As	65
Pb	63
Cd	62
Sb	33
V	30
Ni	24
Be	18
Zn	(10)
B	(10)
F	(10)
Ti	(10)
Cr	nil

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(1) Mainly based on data for Pittsburgh Seam Coal from Ref. 10-10, and indicated at 10% for Zn, B, F and Ti, in the absence of data.

TABLE 10-17

RETENTION OF TRACE ELEMENTS IN GASIFIER ASH

Element	Predicted* Ash Retention, % Present in Coal	Actual Amount in Lurgi Gas, % of Element Present in Coal
Ag	36%	140.1%
As	67	1.11
Be	76	86.0
Hg	4	0.4
Mn	81	134.4
Pb	53	88.5
V	50	114.8
Zn	74	120.6

---

* Assumes 100% of coal ash goes to the gasifier ash.

Slag from the gasifier will be disposed of either by burial in mine spoil or by landfill. The trace elements present in the slag may leach to some extent and enter the ground water. The leaching of trace elements from slag will be lower than leaching from ash. Leaching experiments have been conconducted for power plant ash by Ecology Consultants, Inc., (Ft. Collins, Colorado) and Peabody's Central Laboratory. Both reported that if leaching occurs with water originally meeting drinking water standards, then the leachate will still meet drinking water standards after reaching apparent equilibrium leaching conditions. Specifically, both lead and mercury will occur in this leachate only in the parts per billion range. Similar leaching studies (Reference 10-11) were conducted by Peabody's Central Laboratory on gasifier ash produced with Montana Rosebud Coal and Illinois No. 5 and 6 seam coals. Analyses of leachate from these experiments are presented in Table 10-18.

The raw gas from the gasifier contains entrained char particles. The sizes of some of these could range to micron and submicron levels. Particles in the range of 0.5 to 10 microns in diameter cannot be removed even with high-efficiency cyclones. Concentrations of lead, cadmium, selenium, arsenic, nickel, and chromium increase markedly with a decrease in particle size. Most of these elements are concentrated in particles between 0.5 and 10 μ in diameter. The char particles recovered in the cyclones are generally recycled back to the gasifier.

The raw gas, when scrubbed with water, will have more of the particles removed; a trace quantity of particulates may remain in the gas. Volatilized trace elements present as a vapor will condense and enter the quench water. Table 10-19 (Reference 10-5) shows trace elements in Synthane condensate from an Illinois No. 6 coal gasification test and Table 10-20 (Reference 10-12) shows a breakdown of trace elements found in streams from one commercial Lurgi coal gasification facility. The raw gas from the Lurgi gasifier is scrubbed with water. This mixture is then treated in an oil-water separator. The bottoms from the separator are removed as tar. The top floating layer is tarry oil and is drained out. In the remaining water, tarry gas liquor is further treated to remove phenols and ammonia. Assuming that all the volatilized trace elements end up in the sour water generated by scrubbing the raw gas, and using the volatility data of Table 10-16, the quantities of trace elements in slag and sour water can be estimated. An estimate of the trace element distribution for a BCR gasifier processing coal for a 1,000-MW power plant is shown in Table 10-21.

In the BCR/Air-Blown/Selexol, BCR/Oxygen/Selexol, and IGT/Air-Blown/Selexol processes, trace elements that may be neither condensed during raw gas cooling nor removed during the subsequent scrubbing operation, but remain in the process gas stream may be removed in the acid gas removal unit. The recovery of ammonia from sour water by stripping is done at a temperature of about 230F. Most trace elements present in the sour water will not volatilize at this temperature and, thus, will remain in the water. Part of this water is recycled for scrubbing and a buildup of trace elements may occur. The rest of the stripped water is used as makeup to cooling towers. Trace elements entering with this makeup cooling water may leave by drift or blowdown.

TABLE 10-18

CONTENT OF LEACHATE FROM GASIFIER ASH
(MONTANA ROSEBUD COAL)

Element	Concentration (ppb) 1 Week	 2 Weeks
Lithium	750	820
Lead	100	100
Vanadium	80	80
Antimony	57	49
Zinc	39	12
Chromium	14	20
Copper	11	20
Manganese	10	30
Silver	8	10
Nickel	< 8	< 8
Cadmium	< 3	< 3
Beryllium	< 3	< 3
Arsenic	< 1	< 1
Mercury	0.1	0.5

TABLE 10-19

TRACE ELEMENTS IN CONDENSATE FROM AN ILLINOIS NO. 6 COAL GASIFICATION TEST

Element	ppm
Calcium	4
Iron	3
Magnesium	2
Aluminum	0.8
	ppb
Selenium	360
Potassium	160
Barium	130
Phosphorus	90
Zinc	60
Manganese	40
Germanium	40
Arsenic	30
Nickel	30
Strontium	30
Tin	20
Copper	20
Columbium	6
Chromium	6
Vanadium	3
Cobalt	2

TABLE 10-20

DISTRIBUTION OF TRACE ELEMENTS FOR LURGI GASIFICATION

Element	Ash%	Tarry Gas Liquor%	Tar%	Tar Oil%
Antimony	50.0	45	3.75	1.25
Arsenic	27.0	66.5	2.25	4.25
Boron	92.0	6.5	1.496	.004
Bromine	10.86	89.0	.14	--
Cadmium	52.0	45.0	.90	2.1
Fluorine	56.0	43.917	.08	.003
Lead	93.6	2.054	4.33	.016
Mercury	50.93	40.30	8.12	.65
Nickel	99.554	0.41	.03	.006
Zinc	100%	--	--	--

TABLE 10-21

POTENTIAL RATE OF TRACE ELEMENTS

Element	Feed lbs/hr	Final Residue lbs/hr	Sour Water lbs/hr
Chlorine	1,610.0	161.0	1,449.0
Mercury	0.084	0.008	0.076
Selenium	9.1	2.4	6.7
Arsenic	16.8	4.4	12.4
Lead	7.7	4.9	2.8
Cadmium	0.62	0.23	0.39
Antimony	0.7	0.47	0.23
Vanadium	11.9	8.3	3.6
Nickel	10.5	8.0	2.5
Beryllium	0.7	0.57	0.13
Zinc	34.3	30.9	3.4
Boron	140.0	126.0	14.0
Fluorine	42.7	38.4	4.3
Titanium	490.0	441.0	49.0
Chromium	10.5	10.5	

The importance of trace elements in system design is apparent from the previous discussion. Unfortunately, the estimated fate of these elements is based on scant available data. However, it does indicate that a number of these elements are potentially troublesome. The appearance of so many trace elements in the most troublesome of both the air and water-land columns of Tables 10-12 and 10-14 indicates that these must not be discharged into the stack nor be discharged in a water stream.

While the actual form of the trace element compounds is largely unknown, it is likely that a water wash will be sufficient to control the air emissions to within allowable limits. The ability of a system using high-temperature cleanup to deal with these elements is subject to question. Further, the "dirty" water will require processing prior to disposal. In all likelihood it will be used as quench water for slag in which case virtually all of the trace elements could be concentrated in one location. By proper attention to leaching problems, this could be the most effective means of dealing with the trace elements.

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SECTION 11

PERFORMANCE AND COST OF INTEGRATED SYSTEMS

INTRODUCTION

The performance and cost of eight integrated power systems are presented. The merits of air- vs oxygen-blown operation for a BCR-type gasifier are compared for both a low- and high-temperature sulfur removal system. The U-Gas fluid bed gasifier was evaluated at two steam feed rates with low-temperature desulfurization and at low steam feed rate with high-temperature cleanup. The molten-salt gasifier has integral desulfurization and only a single configuration was investigated. All fuel processing systems were integrated with a high-performance combined-cycle power generating system. Performance of the resulting systems is summarized in Table 11-1 and generating costs are given in Table 11-2.

The results show a reduced difference between high- and low-temperature systems over that previously estimated. Also, little difference in performance and cost of power generation is seen between the air- and oxygen-blown BCR type systems. Not shown here are the emission estimates. Sulfur emissions from the high-temperatures oxygen-blown system are relatively high (0.73 lb/10^6 Btu). Also, nitrogen oxide control may be difficult because of the high combustion temperature associated with the medium-Btu gas from the oxygen-blown systems.

The improved performance of the air-blown systems with low-temperature cleanup should improve the viability of this concept for future clean power generation.

PERFORMANCE

Discussion of the integration of the various systems is divided into three parts. These are low-temperature cleanup, high temperature cleanup, and the molten salt system. A versatile simulation system was used in estimating performance. It permits schematic changes and parametric variations to be made with little difficulty. It is described in Reference 11-1.

TABLE 11-1

SYSTEM PERFORMANCE SUMMARY

			Low-Temperature Cleanup			High-Temperature Cleanup		
Gasification and Cleanup	U-Gas High Steam	U-Gas Low Steam	BCR-Air	BCR-O_2	Molten Salt	U-Gas Low Stm	BCR-Air	BCR-O_2
Coal feed rate - lb/hr	700,000	700,000	700,000	700,000	700,000	700,000	700,000	700,000
Oxidant/coal ratio	3.013	2.864	2.78	.594	2.806	2.864	2.78	.594
Steam/coal ratio	.557	.197	.144	.598	0	.197	.144	.598
Transport gas/coal ratio	.053	.053	.088	.088	0	.053	.088	.088
Gasifier exit temp-F	1660	1660	1700	1700	1800	1660	1,770	1700
Gasifier jacket heat - Btu/lb coal	686	686	214	180	-	686	214	180
Gasifier press - psia	400	400	400	400	400	400	400	400
Raw gas heating value-Btu/SCF (HHV)	138.8	162	171.2	270.7	151.4	162	171.2	270.7
Cold gas efficiency - %	80.1	81.6	83.0	85.7	80.8	81.6	83.0	85.7
Clean gas temperature - F	944	944	926	926	805	1,000	1,000	1,000
Clean gas heating value - Btu/SCF (HHV)	158.6	169.7	177.9	332.4	154	160.9	170.0	268.0
Utilities								
Boost compressor power - MW	32.6	31.0	27.8	22	35.2	31	27.8	22
Let-down turbine power - MW	-	-	-	28	-	-	-	28
Gasifier & Cleanup - MW	27.9	26.7	29.7	20.5	15.7	13.4	21.2	19.4
Boiler & Cooling tower - MW	16.3	16.9	15.8	14.9	14.2	17.2	16.6	15.6
Misc Plant losses - MW	6.5	6.7	6.6	6.0	5.7	6.8	6.8	6.3
Power System *								
Gas Turbine exhaust temp - F	1194	1189	1187	1181	1191	1193	1190	1193
Stack temperature - F	290	300	315	310	299	303	309	318
Output								
Gas turbine - MW	697	712	725	659	618	730	743	694
Steam turbine - MW	460	478	453	422	407	483	465	440
Total - MW	1157	1190	1178	1081	1025	1213	1208	1134
Net plant output - MW	1074	1109	1098	1045	955	1145	1136	1098
Efficiency - Overall (HHV coal)	.429	.443	.439	.418	.433	.457	.454	.439

* Power system parameters for all systems: Gas turbine inlet temperature - 2600 F pressure ratio - 18:1
Steam cycle - 2400 psi/950 F/950 F reheat condenser - 4 in Hg

TABLE 11-2

OVERALL POWER GENERATION COST SUMMARY

	Low-Temperature Cleanup					High Temperature Cleanup		
Capital Costs - $$10^6$	U-Gas High Steam Selexol	U-Gas Low-Stm Selexol	BCR Air Selexol	BCR Oxygen Selexol	Molten Salt	U-Gas Low-Stm Conoco	BCR Air Conoco	BCR Oxygen Conoco
Power System	399	398	404	418	400	361	377	410
Gasification & Cleanup	264	243	253	307	283	182	220	299
Total Capital Cost	663	641	657	725	683	543	597	709
η (HHV Coal)	.429	.443	.439	.418	.433	.457	.454	.439
Electricity Costs - Mills/kWh								
Owning Cost	18.37	17.76	18.20	20.08	18.92	15.04	16.54	19.64
Power Sys. Oper. Cost	2.27	2.27	2.30	2.38	2.28	2.06	2.15	2.34
G&C Oper. Cost	3.64	3.35	3.49	4.24	3.91	2.51	3.04	4.13
Fuel at $1.00/MMBtu	7.96	7.70	7.77	8.17	7.88	7.47	7.52	7.77
Total Mills/kWh	32.24	31.08	31.76	34.87	32.99	27.08	29.25	33.88

The power system, described in Section 7, combines an advanced gas turbine with a high-performance steam cycle for full utilization of the available fuel energy. The gas turbine has a pressure ratio of 18:1 and a firing temperature of 2600 F. Steam cycle throttle conditions are 2400 psi and 950 F with a single reheat to 950 F. Condenser pressure is 4 in. Hg.

The sensitivity of overall peformance to the careful matching of fuel processing and power system characteristics can be best demonstrated by comparing the estimated efficiency of the BuMines-type gasifier integrated with the high-performance power system to previous performance estimates that used a high (24:1) pressure ratio gas turbine and a simple, nonreheat steam cycle. Performance with the better steam system decreases to 35 percent compared to the 37 percent previously estimated. The major difference between the BuMines-type gasifier and those presented here is the inability to use the sensible heat in the fuel gas to raise steam. Because of this decrease in performance, the BuMines-type gasifier was not included in the current detailed comparisons.

Low-Temperature Systems with Selexol Desulfurization

Flow sheets for the four gasifier/low-temperature system combinations using Selexol desulfurization are shown in Figures 11-1 through 11-4. Material balances for the respective systems are given in Tables 11-3 through 11-6 and utility requirements in Tables 11-7 through 11-10. Schematically, the first three gasifier/cleanup configurations are virtually the same, differing only in the point of steam bleed for the gasifier and in the way in which the waste heat boiler is segmented. The oxygen-blown system differs in that it includes a let-down turbine, air cooler, oxygen plant and oxygen compressor. These replace the air cooler and boost compessor needed in the air-blown system to raise the pressure from compressor discharge to gasifier inlet pressure. The let-down turbine extracts enough power to drive the oxygen compressor when operating with a back pressure of 100 psig, the operating pressure required for the air separation process. For both cost and performance reasons bleed air from the gas turbine compressors was used as a source of compressed air for the oxygen plant. The incremental cost of the gas turbine compressor is very small compared to a unit dedicated only to the oxygen plant. From a performance standpoint, the high efficiencies achieved in the gas turbine compressor outweigh the advantages of intercooling normally associated with a separate compressor.

The fuel gas processing schematic is identical for all four systems. Raw fuel gas is first cooled in a boiler to approximately 1200 F then sent to a fuel gas regenerator. The 1200 F value was selected as the maximum temperature desired for the hot side of the regenerative heat exchanger. Dirty gas leaving the regenerator is cooled further in a feedwater heater and subsequent cooler where most of the water vapor is condensed and removed from the gas stream.

A water wash is used for both ammonia and particulate removal. Nearly 100 percent of the ammonia is removed along with equal molar quantities of CO_2 and H_2S. These tend to enhance ammonia removal but require separation so that the H_2S can be sent to the sulfur recovery unit. A Phosam unit is used

SYSTEM FLOW DIAGRAM– U-GAS/ SELEXOL

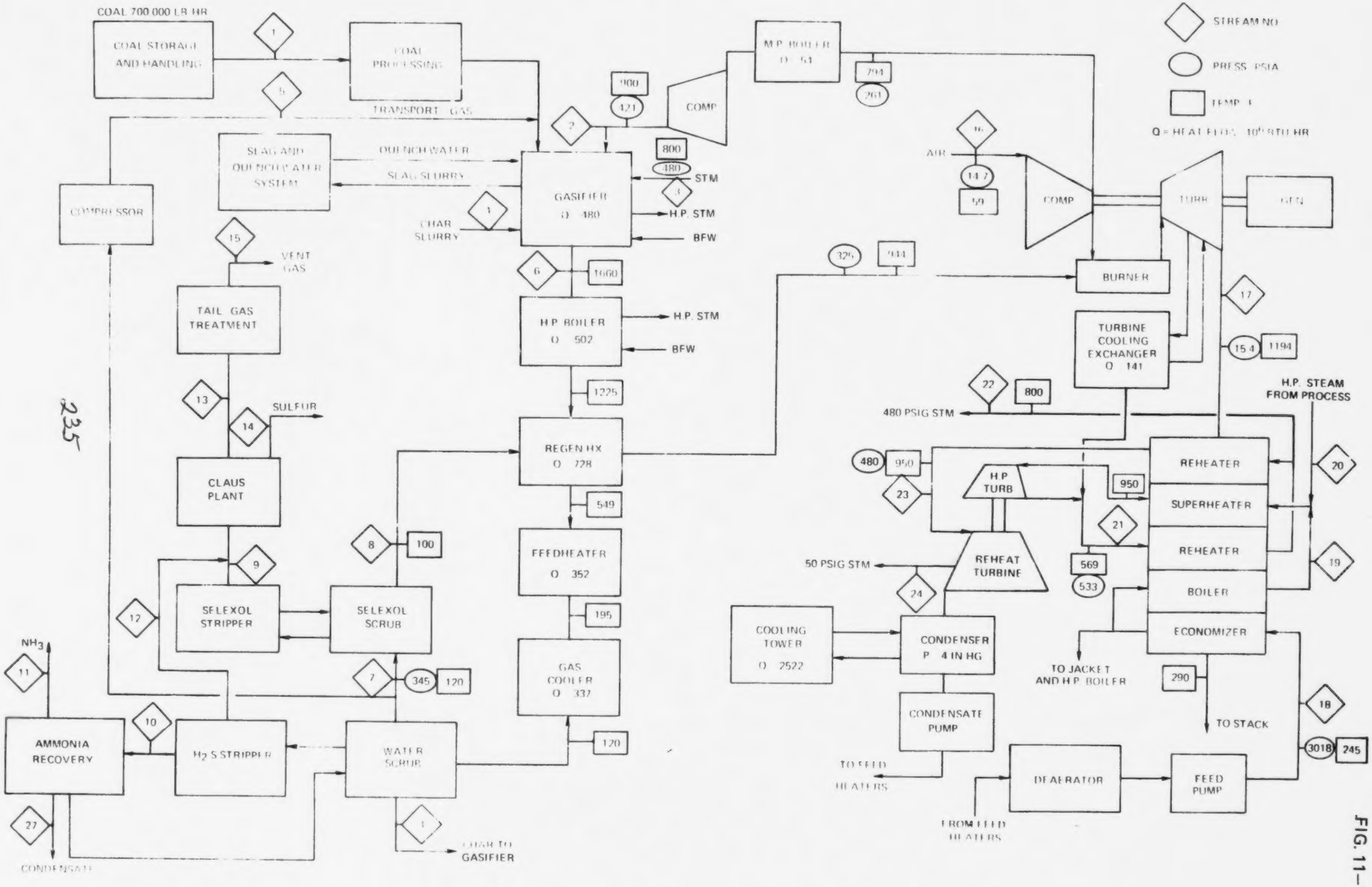
COAL 700,000 LB HR
COAL STORAGE AND HANDLING
COAL PROCESSING
TRANSPORT GAS
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
COMPRESSOR
CHAR SLURRY
GASIFIER Q 480
STM
H.P. STM
BFW
COMP
M.P. BOILER Q 51
H.P. BOILER Q 502
REGEN HX Q 728
FEEDHEATER Q 352
GAS COOLER Q 337
TAIL GAS TREATMENT
VENT GAS
SULFUR
CLAUS PLANT
SELEXOL STRIPPER
SELEXOL SCRUB
WATER SCRUB
H2S STRIPPER
AMMONIA RECOVERY
NH3
CONDENSATE
CHAR TO GASIFIER
AIR
COMP
TURB
GEN
BURNER
TURBINE COOLING EXCHANGER Q 141
480 PSIG STM
50 PSIG STM
H.P. STEAM FROM PROCESS
H P TURB
REHEAT TURBINE
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
TO JACKET AND H.P. BOILER
TO STACK
COOLING TOWER Q 2522
CONDENSER P 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
FROM FEED HEATERS
DEAERATOR
FEED PUMP
STREAM NO
PRESS PSIA
TEMP F
Q = HEAT FLUX 10⁶ BTU HR
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
22
23
24
27
900
121
800
480
794
261
1660
1225
549
195
120
100
345
120
14.7
59
325
944
15.4
1194
800
480
950
950
569
533
290
3018
245

FIG. 11–1

TABLE 11-3

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL

Stream		1		2		3		4	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02			1,592,999	56,852.2				
O_2	32.00			483,603	15,112.6				
H_2O	18.02			32,811	1,820.8	390,029	21,644.2	263	
Coal		700,000							
Char								175	
TOTAL		700,000		2,109,413	73,785.6	390.029	21,644.2	438	

TABLE 11-3 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL

Stream		5		6		7		8	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	820	51.1	64,186	4,001.6	63,369	3,950.7	62,801	3,915.3
H_2	2.016	524	259.7	40,995	20,334.6	40,471	20,075.0	40,410	20,044.5
CO	28.01	8,490	303.1	664,899	23,737.9	656,414	23,435.0	654,235	23,357.2
CO_2	44.01	6,285	142.8	491,944	11,178.0	483,930	10,995.9	406,793	9,243.2
H_2S	34.08	344	10.1	26,562	779.4	26,133	766.8	147	4.3
COS	60.07	18	0.3	1,899	31.6	1,874	31.2	1,226	20.4
NH_3	17.03	--	--	715	42.0	--	--	--	--
N_2	28.02	20,676	737.9	1,619,595	57,801.4	1,598,922	57,063.6	1,595,380	56,937.2
H_2O	18.02	97	5.4	252,507	14,012.6	7,574	420.3	166	9.2
TOTAL		37,254	1,510.4	3,163,302	131,919.1	2,878,687	116,738.5	2,761,158	113,531.1

Stream		9		10		11		12	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	568	35.4						
H_2	2.016	61	30.5						
CO	28.01	2,179	77.8						
CO_2	44.01	77,136	1,752.7					1,730	39.3
H_2S	34.08	25,968	762.5					92	2.7
COS	60.07	649	10.8					--	--
NH_3	17.03	--	--	711	41.8	711.8	41.8	4	0.2
N_2	28.02	3,542	126.4						
H_2O	18.02	7,410	411.2	476,700	26,453.9			2,285	126.8
TOTAL		117,513	3,207.3	477,411	26,495.7	711.8	41.8	4,111	169.0

TABLE 11-3 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL

Stream		13		14		15		16	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	379	23.6			379	23.6		
H_2	2.016	41	20.4			--	--		
CO	28.01	1,451	51.8			1,375	49.1		
CO_2	44.01	80,688	1,833.4			81,110	1,843.0		
H_2S	34.08	279	8.2			--	2 ppm		
COS	60.07	433	7.2			18	0.3		
SO_2	64.06	493	7.7			--	--		
NH_3	17.03	3	0.2			3	0.2		
N_2	28.02	48,802	1,741.7			48,802	1,741.7	10,188,976	361,696
O_2	32.00							3,074,848	96,089
H_2O	18.02	23,940	1,328.5			4,208	233.5		
S	32.06			24,865	775.6				
TOTAL		156,509	5,022.7	24,865	775.6	135,895	3,891.4	13,261,608	457,785

Stream		17		18		19		20	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01	1,608,257	36,543						
CO_2	44.01								
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02	9,997,139	354,886						
O_2	32.00	1,586,880	49,590						
H_2	18.02	502,848	27,905	3,015,000	167,350	0		2,984,785	165,670
S	32.06								
TOTAL		13,695,124	468,924	3,015,000	167,350			2,984,785	165,670

TABLE 11-3 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL

Stream		21		22		23		24	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02								
O_2	32.00								
H_2O	18.02	3,126,316	173,530	396,892	22,030	2,729,424	151,500	144,308	8,010
S	32.06								
TOTAL		3,126,316	173,530	396,892	22,030	2,729,424	151,500	144,308	8,010

SYSTEM FLOW DIAGRAM U–GAS/ LOW STEAM/ SELEXOL

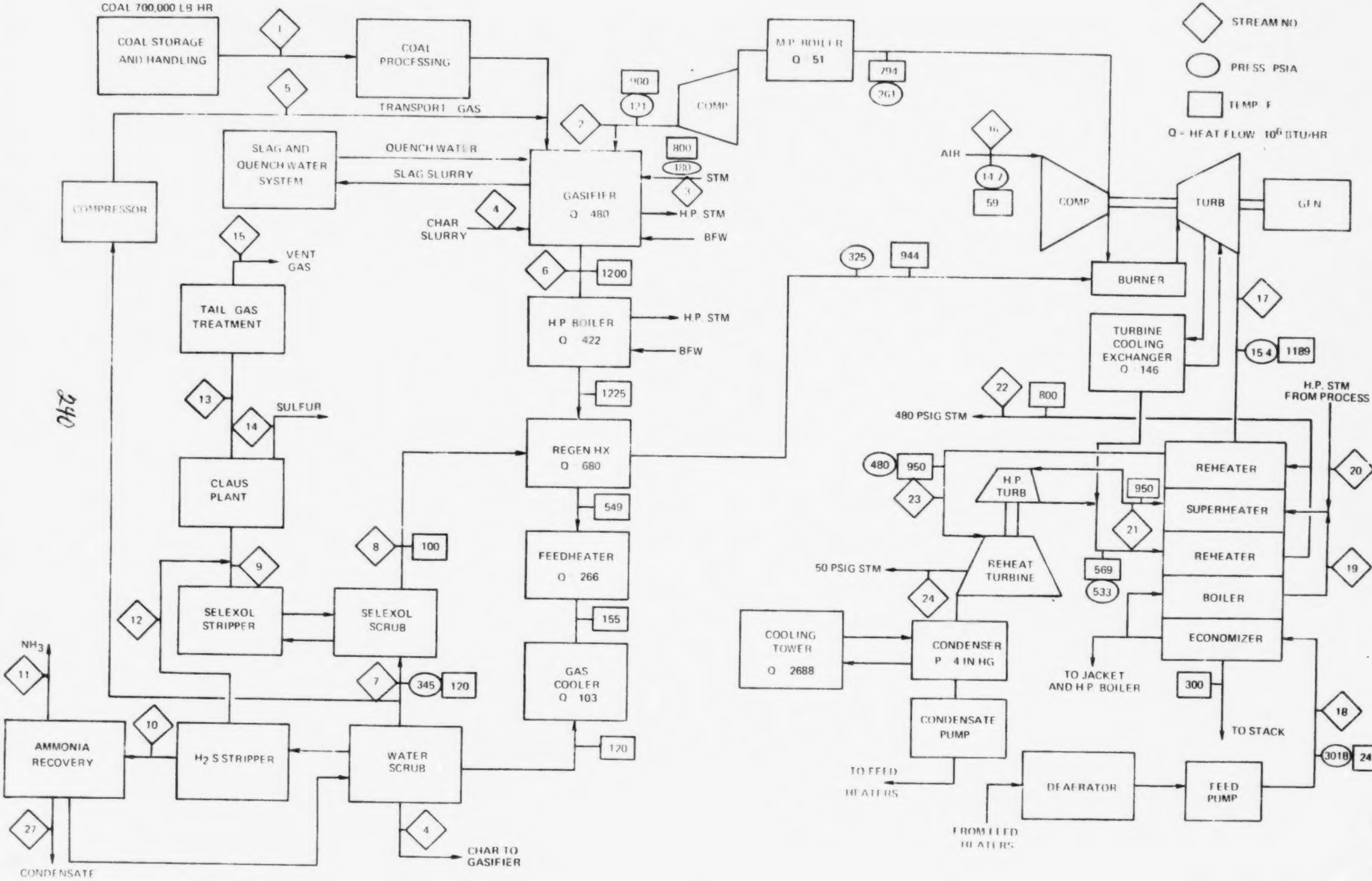
COAL 700,000 LB HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
5
TRANSPORT GAS
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
COMPRESSOR
2
900
121
COMP
M.P. BOILER Q 51
794
261
800
180
3
STM
GASIFIER Q 480
4
CHAR SLURRY
H.P. STM
BFW
6
1200
H.P. BOILER Q 422
H.P. STM
BFW
1225
REGEN HX Q 680
549
FEEDHEATER Q 266
155
GAS COOLER Q 103
120
15
VENT GAS
TAIL GAS TREATMENT
13
14
SULFUR
CLAUS PLANT
9
12
SELEXOL STRIPPER
8
100
SELEXOL SCRUB
7
345
120
NH$_3$
11
AMMONIA RECOVERY
10
H_2S STRIPPER
WATER SCRUB
27
CONDENSATE
4
CHAR TO GASIFIER
STREAM NO
PRESS PSIA
TEMP F
Q = HEAT FLOW 10^6 BTU/HR
16
AIR
14.7
59
COMP
TURB
GEN
325
944
BURNER
17
TURBINE COOLING EXCHANGER Q 146
15.4
1189
H.P. STM FROM PROCESS
22
800
480 PSIG STM
480
950
23
H.P. TURB
950
21
REHEAT TURBINE
50 PSIG STM
24
569
533
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
20
19
COOLING TOWER Q 2688
CONDENSER P 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
TO JACKET AND H.P. BOILER
300
TO STACK
18
3018
245
DEAERATOR
FEED PUMP
FROM FEED HEATERS

FIG. 11–2

TABLE 11-4

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL - LOW STEAM

Stream		1		2		3		4	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02			1,514,078	54.036.6				
O_2	32.00			459,608	14,362.7				
H_2O	18.02			31,215	1,732.2	137,795	7646.8	256	
Coal		700,000							
Char								175	
TOTAL		700,000		2,004,901	70,130.5	137,795	7646.8	431	

TABLE 11-4 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL - LOW STEAM

Stream		5		6		7		8	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	7988	49.8	59,727	3,723.6	58,927.8	3,673.8	58,400	3,640.9
H_2	2.016	456	226.2	34,123	16,926.	33,666.8	16,699.8	33,616	16,674.5
CO	28.01	10,901.5	389.2	815,815	29,125.8	804,912.2	28,736.6	802,240	28,641.2
CO_2	44.01	3,410.8	77.5	256,680	5,832.3	252,018.9	5,726.4	211,847	4,813.6
H_2S	34.08	347.6	10.2	26,008	763.1	25,566.8	750.2	143	4.2
COS	60.07	36.0	0.6	2,489	41.4	2,450.9	40.8	1,604	26.7
NH_3	17.03	--	--	529	31.1	--	--	--	--
N_2	28.02	20,611.5	735.6	1,542,584	55,052.9	1,521,970.7	54,317.3	1,518,560	54,197.0
H_2O	18.02	97.3	5.4	984	3,606.2	7,157.5	397.2	162	9.0
TOTAL		36,660	1,494.5	2,802,939	115,102.4	2,706,670	110,342.1	2,626,572	108,007.1

Stream		9		10		11		12	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	527.7	32.9						
H_2	2.016	51.0	25.3						
CO	28.01	2,672.2	95.4						
CO_2	44.01	40,172.3	912.8					1289.5	29.3
H_2S	34.08	25,423.7	746.0					61.3	1.8
COS	60.07	847.0	14.1						
NH_3	17.03	--	--	526.2	30.9	526.2	30.9	3.4	0.2
N_2	28.02	3,370.8	120.3						
H_2O	18.02	1,679.5	93.2	476,699.3	26,453.9			1693.9	94.0
TOTAL		74,744.2	2,040	477,225.5	26,484.8	526.2	30.9	3048.1	125.3

TABLE 11-4 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL - LOW STEAM

Stream		13		14		15		16	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	351.28	21.9			351.3	21.9		
H_2	2.016	34.1	16.9			--	--		
CO	28.01	1781.4	63.6			1,781.4	63.6		
CO_2	44.01	43,578.7	990.2			44,027.6	1,000.4		
H_2S	34.08	272.6	8.0				2 ppm		
COS	60.07	636.7	10.6			24	0.4		
SO_2	64.06	595.8	9.3			--	--		
NH_3	17.03	3.4	0.2			3	0.2		
N_2	28.02	48,062.7	1715.3			48,062.7	1,715.3	10,606,005	376,500
H_2O	18.02	18,423.6	1022.4			3,222.0	178.8	3,200,640	100,020
S	32.06			24,471.4	763.3				
TOTAL		113,740.28	3,858.4	24,471.4	763.3	97,472.0	2,980.6	13,806,645	476,520

Stream		17		18		19		20	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01	1,632,947	37,104						
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02	10,409,885	369.538						
O_2	32.00	1,722,176	53,818						
H_2O	18.02	431,939	23,970	2,389,035	165,910	204,355	11,343	2,754,773	152,907
TOTAL		14,186,947	484,430	2,989,035	165,910	204,355	11,343	2,754,773	152,907

TABLE 11-4 (Cont'd)

MATERIAL BALANCE FOR U-GAS/AIR-BLOWN/SELEXOL - LOW STEAM

Stream		21		22		23		24	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02								
H_2O	18.02	3,107,189	172,430	144,668	8,030	2,962,488	164,400	136,381	7,570
TOTAL		3,107,189	172,430	144,668	8,030	2,962,488	164,400	136,381	7,570

Stream		25		26		27			
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02								
H_2O	18.02								
TOTAL									

SYSTEM FLOW DIAGRAM BCR/AIR–BLOWN/SELEXOL

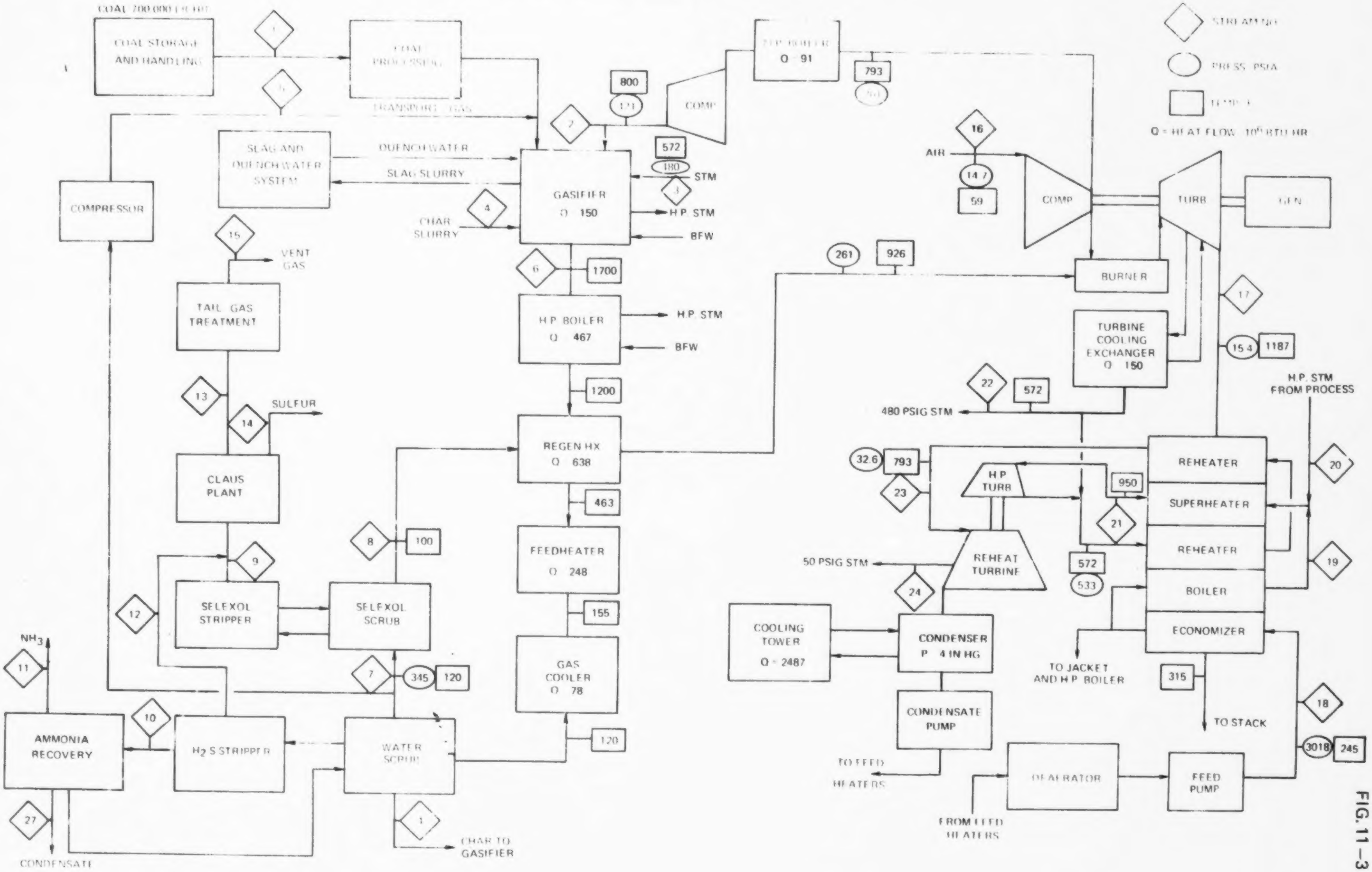
COAL 700,000 LB/HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
5
TRANSPORT GAS
COMPRESSOR
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
2
800
121
COMP
ZIP BOILER Q 91
793
50
572
180
STM
3
GASIFIER Q 150
H.P. STM
BFW
CHAR SLURRY
4
6
1700
H.P. BOILER Q 467
H.P. STM
BFW
1200
REGEN HX Q 638
463
FEEDHEATER Q 248
155
GAS COOLER Q 78
120
15
VENT GAS
TAIL GAS TREATMENT
13
14
SULFUR
CLAUS PLANT
9
SELEXOL STRIPPER
SELEXOL SCRUB
8
100
12
7
345
120
NH3
11
AMMONIA RECOVERY
10
H2 S STRIPPER
WATER SCRUB
1
CHAR TO GASIFIER
27
CONDENSATE
STREAM NO.
PRESS. PSIA
TEMP. F
Q = HEAT FLOW 10^6 BTU/HR
16
AIR
14.7
59
COMP
TURB
GEN
261
926
BURNER
17
TURBINE COOLING EXCHANGER Q 150
15.4
1187
22
572
480 PSIG STM
H.P. STM FROM PROCESS
32.6
793
23
H.P. TURB
950
21
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
20
19
572
533
50 PSIG STM
REHEAT TURBINE
24
COOLING TOWER Q = 2487
CONDENSER P 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
TO JACKET AND H.P. BOILER
315
TO STACK
18
3018
245
DEAERATOR
FEED PUMP
FROM FEED HEATERS

FIG. 11–3

TABLE 11-5

MATERIAL BALANCE FOR BCR/AIR-BLOWN/SELEXOL

Stream		1*		2		3		4	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02			1,469,803	52,455.5				
O_2	32.00			446,208	13,944.0				
H_2O	18.02			10,051	557.8	120,960	6,712.5	4,319	
Coal		700,000							
Char								2,880	
TOTAL		700,000		1,926,062	66,957.3	120,960	6,712.5	7,199	

* Coal is dried to 2 percent moisture prior to firing; flow at the dried condition is 684,286 lb/hr.

TABLE 11-5 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR-BLOWN/SELEXOL

Stream		5		6		7		8	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	1,391	86.7	60,551	3,775.0	59,160	3,688.3	58,631	3,655.3
H_2	2.016	711	352.5	30,875	15,314.9	30,164	14,962.1	30,118	14,939.4
CO	28.01	20,753	740.9	901,631	32,189.6	880,884	31,448.9	877,962	31,344.6
CO_2	44.01	3,446	78.3	149,462	3,396.1	126,273	2,869.2	106,148	2,411.9
H_2S	34.08	596	17.5	25,594	751.0	23,968	703.2	133	3.9
COS	60.07	102	1.7	4,542	75.6	4,440	73.9	2,908	48.4
NH_3	17.03	--	--	8,154	478.8	--	--	--	--
N_2	28.02	34,675	1,237.5	1,506,167	53,753.3	1,471,493	52,515.8	1,468,234	52,399.5
H_2O	18.02	139	7.7	39,871	2,212.6	5,867	325.6	153	8.5
TOTAL		61,813	2,522.8	2,726,847	111,946.9	2,602,249	106,587.0	2,544,287	104,811.5

Stream		9		10		11		12	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	529	33.0						
H_2	2.016	46	22.7						
CO	28.01	2,927	104.5						
CO_2	44.01	20,126	457.3					19,743	448.6
H_2S	34.08	23,836	699.4					1,029	30.2
COS	60.07	1,532	25.5						
NH_3	17.03	--	--	8,104	475.8	8,029	471.4	51	3.0
N_2	28.02	3,259	116.3						
H_2O	18.02	5,714	317.1	374,220	20,766.9			26,048	1,445.5
TOTAL		57,969	1,775.8	382,324	21,242.7	8,029	471.4	46,871	1,927.3

TABLE 11-5 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR-BLOWN/SELEXOL

Stream		13		14		15		16	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	353	22.0			353	22.0		
H_2	2.016	30	15.1			--	--		
CO	28.01	1,949	69.6			1,045	37.3		
CO_2	44.01	42,258	960.2			44,388	1,008.6		
H_2S	34.08	498	14.6			--	2 ppm		
COS	60.07	1,021	17.0			48	0.8		
SO_2	64.06	1,012	15.8			--	--		
NH_3	17.03	32	1.9			32	1.9		
N_2	28.02	47,673	1,701.4			47,673	1,701.4	10,875,282	386,059
O_2	32.00							3,281,920	102,560
H_2O	18.02	45,203	2,508.5			3,179	176.4		
S	32.06			24,178	754.1				
TOTAL		140,029	5,326.1	24,178	754.1	96,718	2,948.4	14,157,202	488,619

Stream		17		18		19		20	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01	1,647,910	37,444						
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02	10,686,064	379,342						
O_2	32.00	1,803,264	56,352						
H_2	18.02	401,432	22,277	2,827,338	156,900	808,917	44,890	1,990,128	110,440
S	32.06								
TOTAL		14,538,670	495,415	2,827,338	156,900	802,115	44,522	1,983,007	110.068

TABLE 11-5 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR-BLOWN/SELEXOL

Stream		21		22		23		24	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02								
O_2	32.00								
H_2O	18.02	2,750,392	152,630	199,661	11,080	2,750,392	152,630	202,500	11,240*
S^2	32.06								
TOTAL									

* Includes steam for deaerator

FIG. 11–4

SYSTEM FLOW DIAGRAM BCR/OXYGEN–BLOWN/ SELEXOL

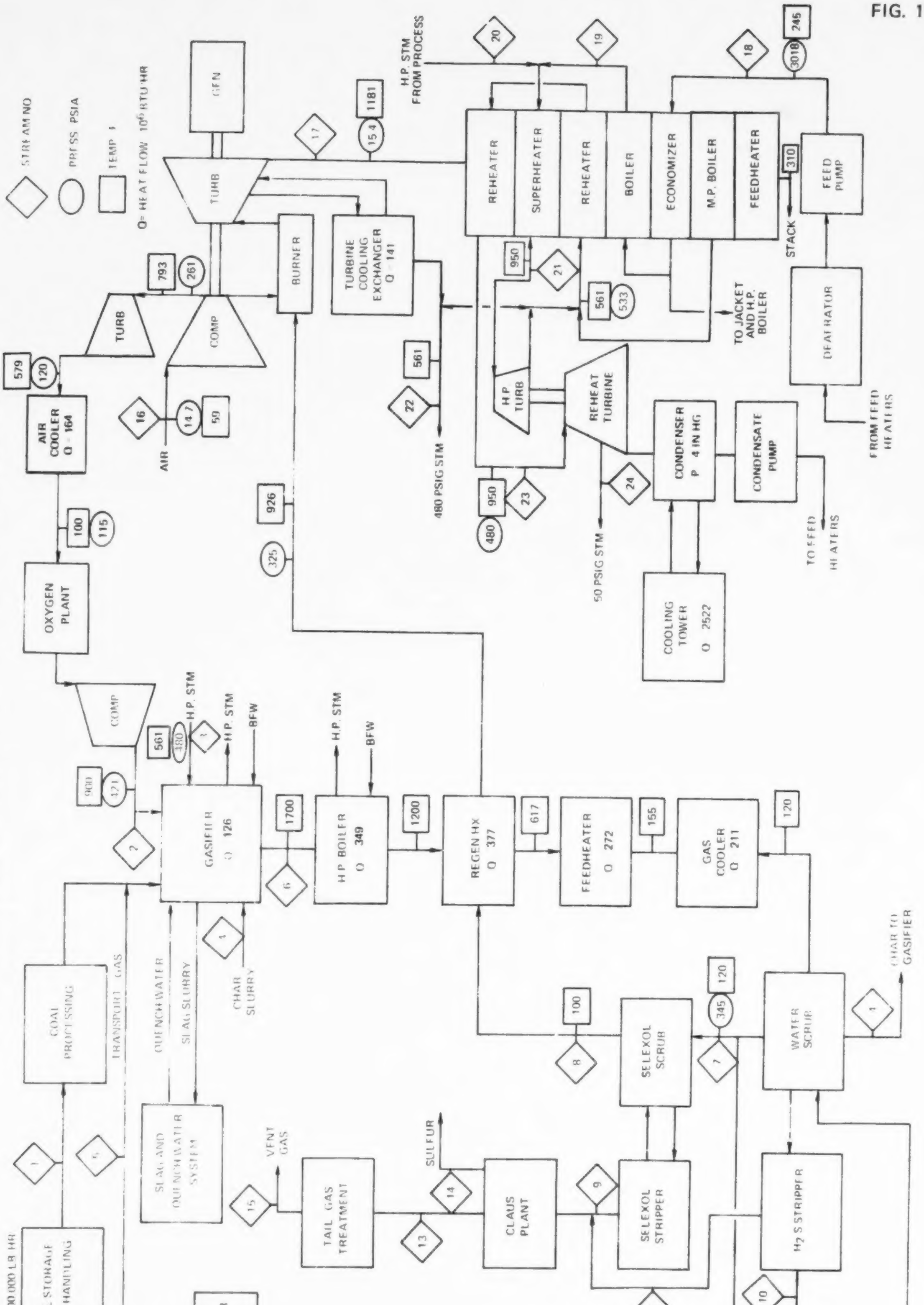
STREAM NO
PRESS PSIA
TEMP F
Q= HEAT FLOW 10⁶ BTU HR
COAL 700,000 LB HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
5
TRANSPORT GAS
QUENCH WATER
SLAG SLURRY
SLAG AND QUENCH WATER SYSTEM
CHAR SLURRY
4
2
900
121
COMP
OXYGEN PLANT
100
115
AIR COOLER Q = 164
579
120
TURB
COMP
793
261
AIR
16
14.7
59
BURNER
GEN
TURB
17
15.4
1181
TURBINE COOLING EXCHANGER Q = 141
926
325
561
22
480 PSIG STM
561
480
3
H.P. STM
BFW
GASIFIER Q 126
1700
6
H.P. BOILER Q 349
H.P. STM
BFW
1200
REGEN HX Q 377
617
FEEDHEATER Q 272
155
GAS COOLER Q 211
120
H.P. STM FROM PROCESS
20
19
18
3018
245
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
M.P. BOILER
FEEDHEATER
310
FEED PUMP
STACK
950
21
561
533
TO JACKET AND H.P. BOILER
DEAERATOR
FROM FEED HEATERS
H.P TURB
REHEAT TURBINE
480
950
23
50 PSIG STM
24
CONDENSER P 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
COOLING TOWER Q 2522
VENT GAS
15
TAIL GAS TREATMENT
SULFUR
14
13
CLAUS PLANT
9
SELEXOL STRIPPER
12
SELEXOL SCRUB
100
8
7
345
120
WATER SCRUB
4
CHAR TO GASIFIER
H₂S STRIPPER
10
COMPRESSOR
AMMONIA RECOVERY
NH₃
11
27
CONDENSATE

TABLE 11-6

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/SELEXOL

Stream		1*		2		3		4	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02			7,302	260.6				
O_2	32.00			408,627	12,769.6				
H_2O	18.02					418,320	23,214.2	3,667	
Coal		700,000							
Char								2,445	
TOTAL		700,000		415,929	13,030.2	418,320	23,214.2	6,112	

* Coal is dried to 2 percent moisture prior to firing; flow at the dried condition is 684,286 lb/hr.

TABLE 11-6 (Cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/SELEXOL

Stream		5		6		7		8	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	3,407	212.4	72,531	4,521.9	69,120	4,309.2	68,500	4,270.6
H_2	2.016	2,093	1,038.2	44,549	22,097.6	42,456	21,059.4	42,391	21,027.4
CO	28.01	34,814	1,242.9	741,108	26,458.7	706,297	25,215.9	703,947	25,132.0
CO_2	44.01	19,364	440.0	412,356	9,369.6	372,276	8,458.8	312,933	7,100.5
H_2S	34.08	1,237	36.3	26,255	770.4	23,938	702.4	133	3.9
COS	60.07	210	3.5	4,620	76.9	4,404	73.3	2,884	48.0
NH_3	17.03	--	--	8,561	502.7	--	--	--	--
N_2	28.02	446	15.9	9,426	336.4	8,980	320.5	8,961	319.8
H_2O	18.02	191	10.6	194,380	10,786.9	3,896	216.2	85	4.7
TOTAL		61,762	2,999.8	1,513,786	74,921.1	1,231,367	60,355.7	1,139,834	57,916.9

Stream		9		10		11		12	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	619	38.6						
H_2	2.016	65	32.0						
CO	28.01	2,350	83.9						
CO_2	44.01	59,339	1,348.3					20,720	470.8
H_2S	34.08	23,805	698.5					1,080	31.7
COS	60.07	1,520	25.3						
NH_3	17.03	--	--	8,508	499.6	8,455	496.5	53	3.1
N_2	28.02	20	0.7						
H_2O	18.02	3,811	211.5	266,700	14,800.2			27,338	1,517.1
TOTAL		91,529	2,438.8	275,208	15,299.8	8,455	496.5	49,191	2,022.7

TABLE 11-6 (Cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/SELEXOL

Stream		13		14		15		16	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	414	25.8			414	25.8		
H_2	2.016	43	21.3			--	--		
CO	28.01	1,566	55.9			852	30.4		
CO_2	44.01	82,224	1,868.3			84,055	1,909.9		
H_2S	34.08	498	14.6			--	2 ppm		
COS	60.07	1,015	16.9			48	0.8		
SO_2	64.06	999	15.6			--	--		
NH_3	17.03	36	2.1			36	2.1		
N_2	28.02	44,644	1,593.3			44,633	1,592.9	11,930,663	425,791
O_2	32.00							3,619,840	113,120
H_2O	18.02	44,726	2,482.0			4,087	226.8		
S	32.06			24,192	754.6				
TOTAL		176,165	6,095.8	24,192	754.6	134,125	3,788.7	15,550,503	538,911

Stream		17		18		19		20	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.0								
H_2	2.016								
CO	28.01								
CO_2	44.01	1,608,389	36,546						
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02	10,387,630	370,722						
O_2	32.00	2,135,328	66,729						
H_2	18.02	533,554	29,609	2,877,073	159,660	1,106,428	61,400	1,527,915	84,790
S	32.06								
TOTAL		13,056,512	467,060	2,877,073	159,660	1,106,428	61,400	1,527,915	84,790

TABLE 11-6 (Cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/SELEXOL

Stream		21		22		23		24	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01								
H_2S	34.08								
COS	60.07								
SO_2	64.06								
NH_3	17.03								
N_2	28.02								
O_2	32.00								
H_2O	18.02	2,517,935	139, 730	496,992	27,580	2,517,935	139,730	61,448	3,410
S	32.06								
TOTAL		2,517,935	139, 730	496,992	27,580	2,517,935	139,730	61,448	3,410

TABLE 11-7

UTILITIES - U-GAS/SELEXOL/AIR-BLOWN

	Coal Preparation	Gasification	Gas Scrubbing & Gas Cooling	Acid Gas Removal	Ammonia Recovery	Sulfur Recovery	Total
Steam #/hr							
420 psig (sat)		390,030			6,900		396,930
100 psig (sat)							
50 psig (sat)				143,300	55,700	(54,600)	144,400
Power - kW	2,500	810	225	22,765	575	1,060	27,935
Cooling Water - gpm			32,680	30,240	805	125	34,730
Cooling Duty MMBtu/hr			326.8	302.4	805	125	638.5

TABLE 11-8

UTILITIES - U-GAS/SELEXOL/AIR-BLOWN - LOW STEAM

	Coal Preparation	Gasification	Gas Scrubbing & Cooling	Acid Gas Removal	Ammonia Recovery	Sulfur Recovery	Total
Steam #/hr							
420 psig (sat)		137,795			6,900		144,695
100 psig (sat)				135,400	55,700	(54,600)	136,500
50 psig (sat)							
Power - kW	2,500	810	225	21,500	575	1,060	26,670
Cooling Water - gpm			10,300	28,600	805	125	33,090
Cooling Duty MMBtu/hr			103	286.0	8	1.3	434.9

TABLE 11-9

UTILITIES BCR/SELEXOL - AIR-BLOWN

Coal Feed - 700,000 #/hr Ill. #6

	Coal Preparation	Gasification	Gas Scrubbing & Cooling	Acid Gas Removal	Ammonia Recovery	Sulfur Recovery	Total
Steam #/hr							
420 psig (sat)		121,000			78,700		199,700
50 psig (sat)				131,500	54,000	(54,600)	130,900
Power - kW	4,360	1,960	80	20,400	1,850	1,060	29,710
Cooling Water - gpm		5,480	7,800	27,770	650	125	36,560
Cooling Duty MMBtu/hr			78	277.7	6.5	1.3	347.7

TABLE 11-10

UTILITIES BCR/SELEXOL - OXYGEN-BLOWN

	Coal Preparation	Gasification	Gas Scrubbing & Cooling	Acid Gas Removal	Ammonia Recovery	Sulfur Recovery	Total
Steam #/hr							
420 psig (sat)		418,320			78,700		497,020
50 psig (sat)				74,800	41,200	(54,600)	61,400
Power - kW	4,360	1,260	80	11,930	1,850	1,060	20,540
Cooling Water - gpm		5,480	21,100	15,800	290	125	24,590
Cooling Duty MMBtu/hr			211	158	2.9	1.3	373.3

to separate the H_2S and recover the ammonia for possible use as a byproduct. Both low- and reheat-pressure steam are required by the Phosam process (see Appendix A). A concentrated slurry of char recovered from the particulate scrubber is returned to the gasifier and recycled to extinction.

Gas leaving the water wash is sent to the Selexol acid-gas removal system. Operation of that unit is also described in Section 6. Steam bled from the power system is used in the stripper. Acid gas from the Selexol stripper varies between 24 and 40 percent in H_2S concentration, the particular value being a direct function of the amount of CO_2 in the fuel gas. A small percentage of the CO_2 (16 percent) is absorbed and finds its way into the acid gas. To this stream is added the H_2S recovered from the water scrub and the combined total is sent to the Claus plant for sulfur recovery. Low-pressure steam is raised in the burner and sulfur condensers of that process. The operational details of the Claus plant and Beavon tail gas cleanup process are as given in Section 6. Nearly all of the sulfur is recovered and emissions from this part of the process are relatively insignificant.

The clean fuel gas exiting the Selexol process is reheated to approximately 925 F by heat exchange with the raw gas and delivered to the gas turbine fuel control.

The major source of heat for the steam cycle is the waste heat recovery boiler, although some steam is raised at reheat pressure by heat exchange with the turbine cooling air. In this case, boiler is a misnomer in that virtually all the steam is raised in the gasifier jacket and/or the heat recovery boiler in the gasifier exit stream.

The reheater is shown in two sections. This minimizes the pinch problem that would occur in the downstream unit of a series of combinations of reheater and superheater. An alternative is to arrange them in parallel in the gas flowstream. In this arrangement, reheat-pressure stream is available at bleed temperature, reheat temperature or an intermediate temperature as required for the gasifier. In the air-blown systems, very little, if any, steam is raised in the waste heat boiler in the gas turbine exhaust. However, in the oxygen-blown system, the smaller fuel gas flow rate results in reduced quantities of heat and, therefore, less steam is available from the waste heat boiler. In this system, it is necessary to raise both high- and reheat-pressure steam in the gas turbine exhaust waste heat boiler. This is done in order to utilize all of the available heat in the exhaust stream. As a result, temperature differences are as small as practical, making for a larger and more complex heat exchanger. This shows up in steam cycle cost for the oxygen-blown system. (It is possible that slight modifications in the steam cycle could reduce the size of the waste heat boiler but the changes would not materially affect study conclusions.)

The steam turbine is conventional in all respects except for bleed provisions. Because of the large amount of waste heat available from the process and gas turbine exhaust, bleed for feedwater heating is unnecessary. The condenser pressure of 4 in. Hg is compatible with cooling tower use. In general, the combined cycle is not very sensitive to condenser pressure and a higher level could be helpful if it were desired to reduce water use rate in the cooling tower.

As expected, the two air-blown systems with low steam feed rates had similar (.443 and .439) efficiencies. The U-Gas system with high steam feed was over one point less and the oxygen-blown system approximately two points less.

Molten Salt System

With the exception of the fuel processing system, operation is quite similar to the other low-temperature systems. Both gasifier and fuel processing systems are described in Section 6. The system flow diagram is shown in Figure 11-5 the material balance in Table 11-11 and utility requirements in Table 11-12.

The molten salt gasifier does not require steam feed. However, the salt recovery process needs 200 psi steam for calcination of the sodium bicarbonate. Also, some low-pressure steam is used to preheat the sodium carbonate solution prior to entering the CO_2-H_2S absorber. This steam is raised in the Claus plant. Also, part of the calciner steam requirement is raised in the Claus plant. The remainder is raised by heat exchange with the turbine cooling air and superheated in the waste heat boiler. Some of that steam is used for the feedwater pump drive.

A further difference between the molten salt and the other low-temperature systems results from the fact that the fuel gas is not cooled below its dew point. In the Selexol systems, the lower heat capacity of the dry gas on the cold side of the regenerative heat exchanger results in an elevated exit temperature on the hot side with the remainder of the sensible heat used for feedwater heating. In this system, it was necessary to use some of the heat from the gas turbine exhaust waste heat boiler for feedwater heating. This increased the temperature differences in this waste heat boiler thereby reducing boiler size and cost. Quite possibly, a slight improvement in efficiency could be achieved by the use of regenerative feedwater heating up to 250 F. However, it would not materially affect the results or conclusions of the study.

High-Temperature Systems

The three systems using Conoco desulfurization are shown in Figures 11-6, 11-7, and 11-8. Material balances are given in Tables 11-13, 11-14, and 11-15 and utility requirements in Tables 11-16, 11-17, and 11-18. The same differences between air- and oxygen-blown systems appear here as in the low-temperature systems. While all systems required the use of the gas turbine exhaust waste heat boiler for feedwater heating, the oxygen-blown design again proved more difficult and a stack temperature lower than 318 F was not achievable.

The fuel gas flow path for these systems is quite simple. Raw gas from the gasifier goes directly to the desulfurizer except in the air-blown BCR case. There, the desulfurizer cannot operate above 1600 F due to the low CO_2 partial pressure in the gas stream. (The temperature limitation of 1000 F is to prevent calcining of the absorbant during desulfurization.) In that system

MOLTEN SALT SYSTEM FLOW DIAGRAM

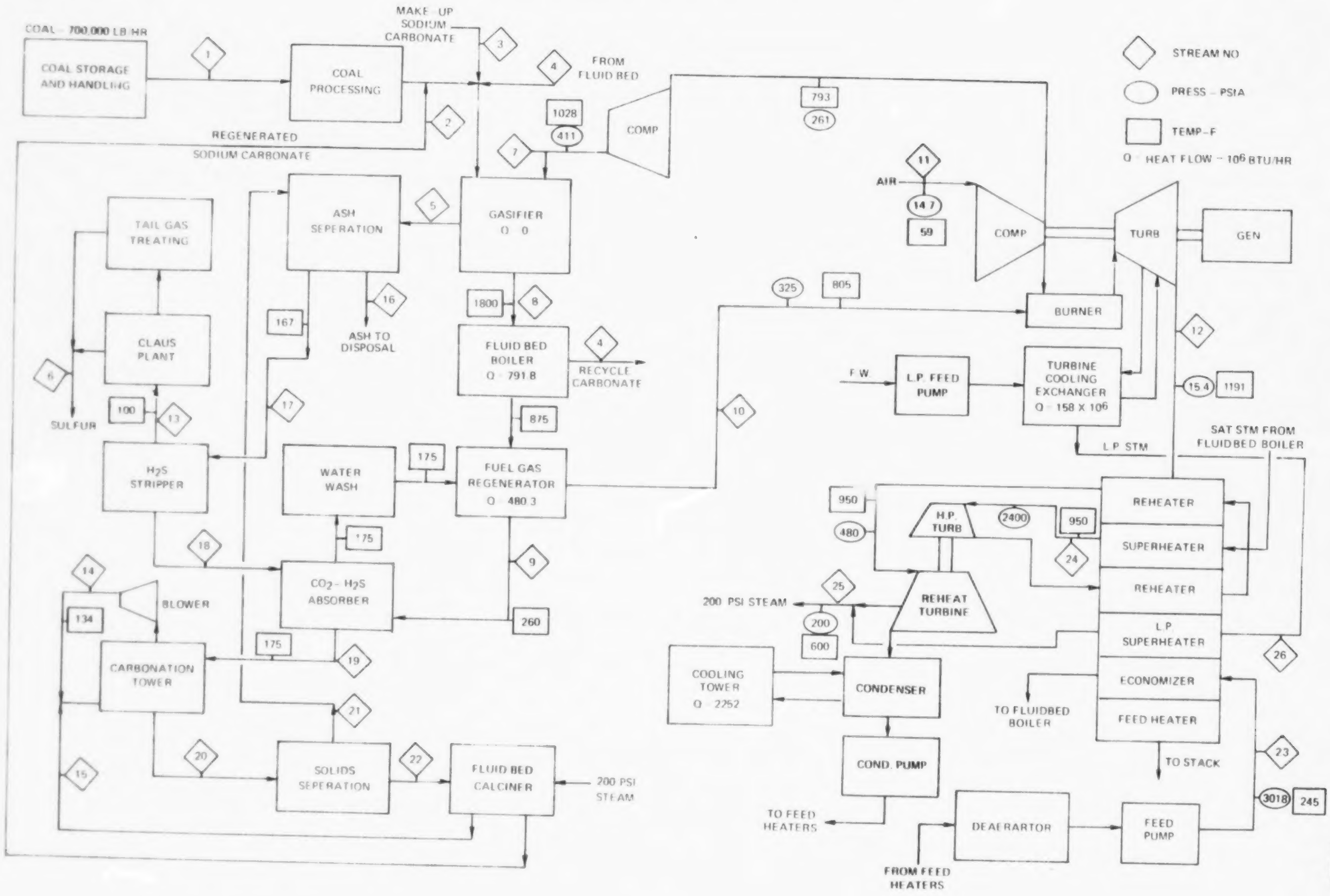
STREAM NO
PRESS – PSIA
TEMP–F
Q – HEAT FLOW – 10⁶ BTU/HR
COAL – 700,000 LB/HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
MAKE–UP SODIUM CARBONATE
3
4
FROM FLUID BED
2
REGENERATED SODIUM CARBONATE
1028
411
7
COMP
793
261
GASIFIER Q – 0
ASH SEPERATION
5
16
ASH TO DISPOSAL
167
TAIL GAS TREATING
CLAUS PLANT
6
SULFUR
100
13
H2S STRIPPER
17
1800
8
FLUID BED BOILER Q – 791.8
4
RECYCLE CARBONATE
875
WATER WASH
175
FUEL GAS REGENERATOR Q – 480.3
10
325
805
18
CO2 – H2S ABSORBER
175
9
260
14
BLOWER
134
CARBONATION TOWER
175
19
21
20
SOLIDS SEPERATION
22
FLUID BED CALCINER
200 PSI STEAM
15
AIR
11
14.7
59
COMP
TURB
GEN
BURNER
12
F.W.
L.P. FEED PUMP
TURBINE COOLING EXCHANGER Q – 158 X 10⁶
15.4
1191
L.P. STM
SAT STM FROM FLUIDBED BOILER
950
480
H.P. TURB
2400
950
24
REHEATER
SUPERHEATER
REHEATER
L.P. SUPERHEATER
ECONOMIZER
FEED HEATER
26
200 PSI STEAM
25
200
600
REHEAT TURBINE
COOLING TOWER Q – 2252
CONDENSER
COND. PUMP
TO FEED HEATERS
TO FLUIDBED BOILER
TO STACK
23
3018
245
DEAERARTOR
FEED PUMP
FROM FEED HEATERS

FIG. 11–5

TABLE 11-11

MOLTEN SALT SYSTEM

Stream	1	2	3	4	5	6
Component	LB/HR	LB/HR	LB/HR	LB/HR	LB/HR	LB/HR
Coal - as rec'd	700,000					
Na CO		169,565	30,730	2,606	128,682	
Na S		482			44,548	
NaOH					8,885	
Ash					72,590	
Sand				64	64	
Sulfur						19,616
Total Solids and Melt	700,000	170,047	30,730	2,670	254,769	19,616

TABLE 11-11 (Cont'd)

MOLTEN SALT SYSTEM

Stream		7		8		9		10	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04			25,585.4	1,595.1	25,585.4	1,595.1	25,585.4	1,595.1
H_2	2.016			29,333.0	14,649.3	29,533.0	14,649.3	29,533.0	14,649.3
CO	28.01			843,101.	30,100.0	843,101.0	30,100.0	843,101.0	30,100.0
CO_2	44.01			144,040.3	3,272.9	144,040.3	3,272.9	101,487.1	2,306.0
H_2S	34.08			3,558.0	104.4	3,558.0	104.4	603.2	17.7
COS	60.07			426.5	7.1	426.5	7.1	426.5	7.1
SO_2	64.06								
N_2	28.02	1,506,988	53,782.6	1,513,968.2	54,031.7	1,513,968.2	54,031.7	1,513,968.2	54,031.7
O_2	32.0	457,504	14,297.						
H_2O	18.02			44,958.1	2,494.9	44,958.1	2,494.9	29,365.4	1,629.6
Na_2CO_3	105.99			2,607.4	24.6				
TOTAL		1,964,492	68,079.6	2,607,770.9	106,280.0	2,605,170.5	106,255.4	2,544,069.8	104,336.5

Stream		11		12		13		14	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01			1,468,129.6	33,768.	11,072.9	251.6	1,064,865.9	24,196.
H_2S	34.08					20,891.	613.0	35,170.6	1,032.
COS	60.07								
SO_2	64.06			1,588.7	24.8				
N_2	28.02	9,157,269	326,812	9,006,048.3	321,415.	350.25	12.5	64,530.1	2,303.0
O_2	32.0	2,780,040	86,876	1,448,960.0	45,280.	105.6	3.3	19,590.4	612.2
H_2O	18.02			365,247.4	20,269.	17,823.6	989.1	1,817,497.2	1,008.6
Na_2CO_3	105.99								
TOTAL		11,937,309	413,688.2	12,307,973.	420,756.8	50,243.35	1,869.5	1,202,331.9	29,151.8

TABLE 11-11 (Cont'd)

MOLTEN SALT SYSTEM

Stream		15	
Comp	M.W.	LB/HR	MOLS/HR
CH_4	16.04		
H_2	2.016		
CO	28.01		
CO_2	44.01	68,888.9	1,565.3
H_2S	34.08	211.3	6.2
COS	60.07		
SO_2	64.06		
N_2	28.02	350.25	12.5
O_2	32.0	105.6	3.3
H_2O	18.02		
Na_2CO_3	105.99		
TOTAL		69,556.05	1,587.3

Stream		16		17		18		19	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
Na_2CO_3	105.99	4,840	45.67	244,303	2,304.9	321,422	3,032.6	209,747	1,978.9
$NaHCO_3$	84.01	9,873	117.5	136,922	1,629.83	65,696	782	235,439	2,802.5
NaHS	56.06	2,150	38.35	66,868	1,192.79	32,817	585.4	37,677	672
H_2O	18.02	81,388	4,516.5	1,814,112	100,672.14	1,760,601	97,702.6	1,758,770	97,600.9
TOTAL		98,251	4,718.02	2,262,205	105,799.66	2,180,536	102,102.6	2,241,633	107,835.7

TABLE 11-11 (Cont'd)

MOLTEN SALT SYSTEM

Stream		20		21		22		23	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
Na_2CO_3	105.99	57,716	544.5	56,656	534.5	1,060	10.00		
$NaHCO_3$	84.01	217,397	2,587.8	213,406	2,540.2	3,991	47.50		
NaHS	56.06	37,708	672.6	37,016	660.3	693	12.36		
H_2O	18.02	1,611,046	89,403.2	1,581,471	87,762	29,575	1,641.23	2,428,766	134,781.7
TOTAL		1,923,867	93,208.1	1,888,549	91,497	35,319	1,711.09		

Stream		24		25		26	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
Na_2CO_3	105.99						
$NaHCO_3$	84.01						
NaHS	56.06						
H_2O	18.02	2,404,456	133,432.6	619,927	34,402.2	128,030	7,104.9

TABLE 11-12

UTILITY SUMMARY - MOLTEN SALT SYSTEM

	Coal Handling	Gasification & Gas Cleanup	Ash Removal Salt Recovery	Sulfur Recovery	Total
Steam = lb/hr, 200 psi	-		182,700	(37,800)	144,900
50 psi	-		11,400	(11,400)	-
Power - kW	2,500	100	12,200	850	15,650
Cooling Water - GPM			40,100	120	40,220
Cooling Duty - MMBtu/hr			401	1.2	402
Make-up Na_2CO_3 - TPD		369			369

SYSTEM FLOW DIAGRAM U–GAS/ LOW STEAM/ CONOCO

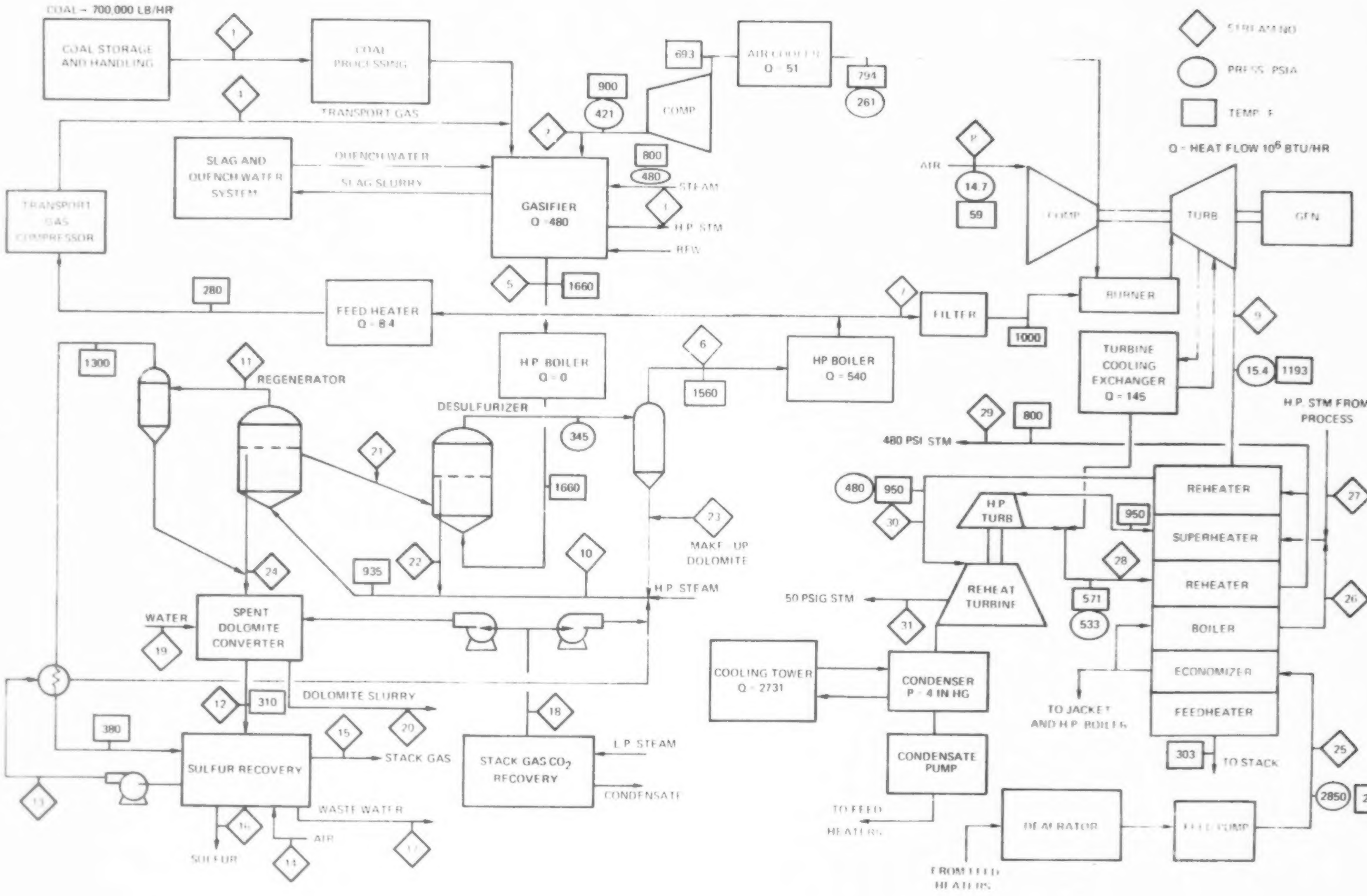
COAL – 700,000 LB/HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
TRANSPORT GAS
TRANSPORT GAS COMPRESSOR
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
GASIFIER Q = 480
2
900
421
COMP
693
AIR COOLER Q = 51
794
261
800
480
STEAM
H P STM
BFW
5
1660
280
FEED HEATER Q = 8.4
H P BOILER Q = 0
7
FILTER
1000
BURNER
AIR
14.7
59
COMP
TURB
GEN
9
TURBINE COOLING EXCHANGER Q = 145
15.4
1193
H.P. STM FROM PROCESS
STREAM NO
PRESS PSIA
TEMP F
Q = HEAT FLOW 10^6 BTU/HR
1300
11
REGENERATOR
DESULFURIZER
345
1660
21
6
1560
HP BOILER Q = 540
23
MAKE UP DOLOMITE
24
935
22
10
H P STEAM
WATER
19
SPENT DOLOMITE CONVERTER
12
310
DOLOMITE SLURRY
20
18
380
SULFUR RECOVERY
15
STACK GAS
STACK GAS CO2 RECOVERY
L P STEAM
CONDENSATE
13
16
SULFUR
14
AIR
WASTE WATER
17
29
800
480 PSI STM
480
950
30
H P TURB
950
28
571
533
REHEAT TURBINE
50 PSIG STM
31
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
FEEDHEATER
27
26
TO JACKET AND H P BOILER
303
TO STACK
25
2850
245
COOLING TOWER Q = 2731
CONDENSER P = 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
DEAERATOR
FEED PUMP
FROM FEED HEATERS

FIG. 11–6

TABLE 11-13

MATERIAL BALANCE FOR U-GAS/LOW STEAM/CONOCO

Stream		1		2		3	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04						
H_2	2.016						
CO	28.01						
CO_2	44.01						
H_2S	34.08						
COS	60.07						
NH_3	17.03						
N_2	28.02			1,514,078	54,035.6		
O_2	32.00			459,608	14,362.7		
H_2O	18.02			31,215	1,732.2	137,795	7,646.8
Coal Char		700,000					
TOTAL		700,000		2,004,901	70,130.5	137,795	7,646.8

TABLE 11-13 (Cont'd)

MATERIAL BALANCE FOR U-GAS/LOW STEAM/CONOCO

Stream		4		5		6		7	
Comp.	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	788	49.1	59,727	3,723.6	59,727	3,723.6	58,939	3,674.5
H_2	2.016	452	224.2	34,123	16,926.0	34,279	17,003.3	33,827	16,779.1
CO	28.01	10,731	383.1	815,815	29,125.8	813,648	29,048.5	802,917	28,665.4
CO_2	44.01	3,908	88.8	256,680	5,832.3	296,275	6,732.0	292,367	6,643.2
H_2S	34.08	7	0.2	26,008	763.1	562	16.5	555	16.3
COS	60.07	--	--	2,489	41.4	210	3.5	210	3.5
NH_3	17.03	7	0.4	529	31.1	530	31.1	523	30.7
N_2	28.02	20,343	726.0	1,542,584	55,052.9	1,542,582	55,052.9	1,522,239	54,326.9
H_2O	18.02	1,016	56.4	64,984	3,606.2	77,045	4,275.5	76,029	4,219.1
TOTAL		37,252	1528.2	2,802,939	115,102.4	2,824,858	115,886.9	2,787,606	114,358.7

Stream		8		9		10		11	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
O_2	32.0	3,172,192	99,131	1,692,768	52,899.0				
N_2	28.02	10,455,607	373,148	10,260,420	366,182.0				
CO_2	44.01			1,715,800	38,986.6	404,822	9,198.4	380,189	8,638.7
H_2O	18.02			511,941	28,409.6	90,403	5,016.8	77,836	4,319.4
SO_2	64.06			1,268	19.8				
NO	30.01			921	30.7				
H_2S						3,050	89.5	26,818	786.9
TOTAL		13,627,799	472,279	14,183,118	486,527.7	498,275	14,304.7	484,843	13,745.0

Stream		12		13		14		15	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CO_2	44.01	2,711	61.6	382,900	8,700.3				
H_2O	18.02	366	20.3	55,363	3,072.3			72	4.0
H_2S	34.08	3,688	108.2	3,050	89.5				
O_2	32.0					14,352	448.5	1,306	40.8
N_2	28.02					47,275	1,687.2	47,275	1,687.2
SO_2	64.06							314	4.9
TOTAL		6,765	190.1	441,313	11,862.1	61,627	2,135.7	48,967	1,736.9

TABLE 11-13 (Cont'd)

MATERIAL BALANCE FOR U-GAS/LOW STEAM/CONOCO

Stream		16		17		18		19	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
H_2O	18.02			37,283	2,069	2,663	147.8	35,135	1,949.8
CO_2	44.01					35,455	805.6		
S	32.06	25,670	800.7						
TOTAL		25,670	800.7	37,283	2,069	38,118	953.4	35,135	1,949.8

Stream		20		21		22		23	
Comp.	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
$MgCO_3 \cdot CaCO_3$	184.1	25,338	137.7					25,338	137.7
$MgO \cdot CaS$	112.46			608,634	5,412.0	699,231	6,217.6		
$MgO \cdot CaCO_3$	140.41			206,894	1,473.5	93,780	667.9		
Inert Solids	100.0	2,340	23.4	117,050	1,170.5	117,050	1,170.5	2,340	23.4
H_2O	18.02	33,836	1,877.7						
TOTAL		61,514	2,038.8	935,578	8,056.0	910,061	8,056.0	27,678	161.1

Stream		24	
Comp.	M.W.	LB/HR	MOLS/HR
$MgO \cdot CaS$	112.46	12,168	108.2
$MgO \cdot CaCO_3$	140.41	4,142	29.5
Inert Solids	100.0		
TOTAL		16,310	137.7

SYSTEM FLOW DIAGRAM– BCR/AIR –BLOWN/ CONOCO

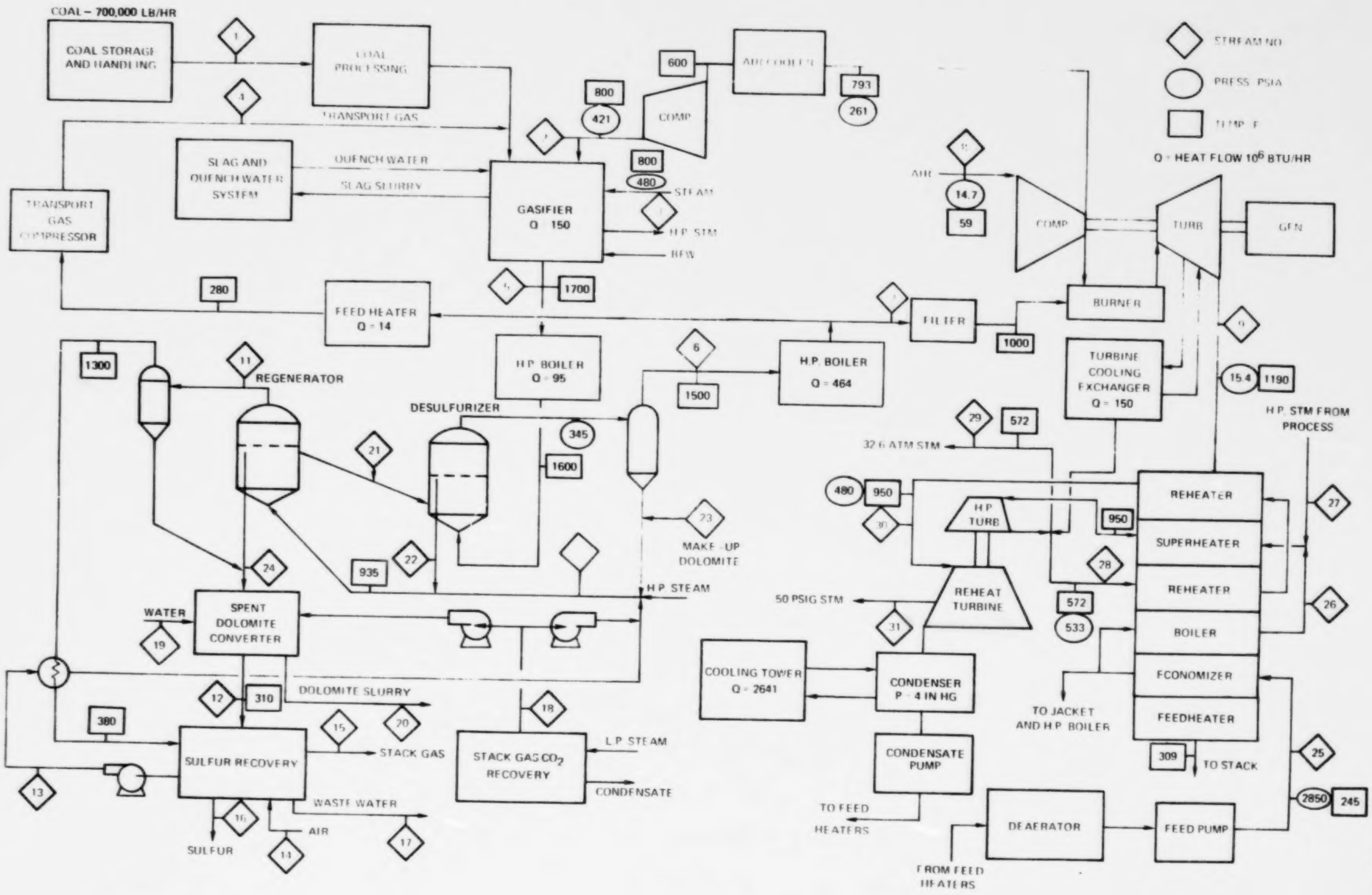
COAL – 700,000 LB/HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
4
TRANSPORT GAS
TRANSPORT GAS COMPRESSOR
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
800
421
COMP
600
AIRCOOLER
793
261
800
480
STEAM
H.P. STM
BFW
GASIFIER Q 150
5
1700
280
FEED HEATER Q = 14
H.P. BOILER Q = 95
6
1500
H.P. BOILER Q = 464
FILTER
1000
BURNER
AIR
8
14.7
59
COMP
TURB
GEN
9
15.4
1190
TURBINE COOLING EXCHANGER Q = 150
H.P. STM FROM PROCESS
STREAM NO
PRESS PSIA
TEMP F
Q = HEAT FLOW 10^6 BTU/HR
1300
11
REGENERATOR
DESULFURIZER
21
345
1600
22
23
MAKE-UP DOLOMITE
935
24
H.P. STEAM
WATER
19
SPENT DOLOMITE CONVERTER
12
310
DOLOMITE SLURRY
20
380
13
SULFUR RECOVERY
15
STACK GAS
16
SULFUR
14
AIR
WASTE WATER
17
18
STACK GAS CO2 RECOVERY
L.P. STEAM
CONDENSATE
29
572
32.6 ATM STM
480
950
30
H.P. TURB
REHEAT TURBINE
950
28
572
533
50 PSIG STM
31
COOLING TOWER Q = 2641
CONDENSER P - 4 IN HG
CONDENSATE PUMP
TO FEED HEATERS
FROM FEED HEATERS
DEAERATOR
FEED PUMP
2850
245
25
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
FEEDHEATER
27
26
TO JACKET AND H.P. BOILER
309
TO STACK

FIG. 11 –7

TABLE 11-14

MATERIAL BALANCE FOR BCR/AIR BLOWN/CONOCO

Stream		1		2		3	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04						
H_2	2.016						
CO	28.01						
CO_2	44.01						
H_2S	34.08						
COS	60.07						
NH_3	17.03						
N_2	28.02			1,469,803	52,455.5		
O_2	32.00			446,208	13,944.0		
H_2O	18.02			10,051	557.8	120,960	6.712.5
Coal		700,000					
Char							
TOTAL		700,000		1,926,062	66,957.3	120,960	6,712.5

TABLE 11-14 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR BLOWN/CONOCO

Stream		4		5		6		7	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	1,355	84.5	60,551	3,775.0	60,551	3,775.0	59.196	3,690.5
H_2	2.016	717	355.6	30,875	15,314.9	32,043	15,894.5	31,326	15,538.9
CO	28.01	19,809	707.2	901,631	32,189.6	885,396	31,610.0	865,587	30,902.8
CO_2	44.01	4,788	108.8	149,462	3,396.1	214,038	4,863.4	209,250	4,754.6
H_2S	34.08	--	--	25,594	751.0	324	9.5	324	9.5
COS	60.07	--	--	4,542	75.6	150	2.5	150	2.5
NH_3	17.03	182	10.7	8,154	478.8	8,154	478.8	7,972	468.1
N_2	28.02	33,697	1,202.6	1,506,167	53,753.3	1,506,167	53,753.3	1,472,471	52,550.7
H_2O	18.02	957	53.1	39,871	2,212.6	42,789	2,374.5	41,832	2,321.4
TOTAL		61,505	2,522.5	2,726,847	111,946.9	2,749,612	112,761.5	2,688,108	110,239.0

Stream		8		9		10		11	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
O_2	32	3,282,240	102,570	1,794,080	56,065	--	--	--	--
N_2	28.02	10,818,214	386,089	10,623,223	37,913	--	--	--	--
CO_2	44.01	--	--	1,733,158	39,381	409,350	9,301.3	384,440	8,735.3
H_2O	18.02	--	--	455,239	25,263	91,414	5,072.9	78,707	4,367.7
SO_2	64.02	--	--	640	10	--	--	--	--
NO	30.01	--	--	--	--				
H_2S	34.08	--	--	--	--	3,084	90.5	27,118	795.7
TOTAL		14,100454	488,659	14,606,340	499,849	503,848	23,766.0	490,265	13,898.7

TABLE 11-14 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR BLOWN/CONOCO

Stream		12		13		14		15	
Comp	M. W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CO_2	44.01	2,742	62.3	387,182	8,797.6	--	—	—	—
H_2O	18.02	371	20.6	55,981	3,106.6	--	—	72	4.0
H_2S	34.08	3,728	109.4	3,084	90.5	--	—	—	—
O_2	32.0	--	--	--	--	14,509	453.4	132	41.2
N_2	28.02	--	--	--	--	47,791	1,705.6	47,791	1,705.6
SO_2	64.06	--	--	--	--	--	--	314	4.9
TOTAL		6,841	192.3	446,247	11,994.7	62,300	2,159.0	48,309	1,755.7

Stream		16		17		18		19	
Comp	M. W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
H_2O	18.02	--	--	37,703	2,092.3	2,692	149.4	35,526	1,971.5
CO_2	44.01	--	--	--	--	35,851	814.6	—	—
S	32.06	25,959	809.7	--	--	--	--	—	—
TOTAL		25,959	809.7	37,703	2,092.3	38,543	964.0	35,526	1,971.5

TABLE 11-14 (Cont'd)

MATERIAL BALANCE FOR BCR/AIR BLOWN/CONOCO

Stream		20		21		22		23	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
$MgCo_3 \cdot CaCO_3$	184.01	25,614	139.2	--	--	--	--	25,614	139.2
$MgO \cdot CaS$	112.46	--	--	612,300	5,444.6	703,910	6,259.2	--	--
$MgO \cdot CaCo_3$	140.41	--	--	213,114	1,507.8	98,736	703.2	--	--
Inert Solids	100.0	2,370	23.7	118,360	1,183.6	118,360	1,183.6	2,370	23.7
H_2O	18.02	34,211	1,898.5	943,774	--	--	--	--	--
TOTAL		62,195	2,061.4	1,887,548	8,146.0	921,006	8,146.0	27,984	162.9

Stream		24							
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
$MgO \cdot CaS$	112.16	12,303	109.4						
$MgO \cdot CaCo_3$	140.41	4,184	29.8						
Inert Solids	100.0	2,370	23.7						
TOTAL		18,857	162.9						

SYSTEM FLOW DIAGRAM– BCR/ OXYGEN–BLOWN/ CONOCO

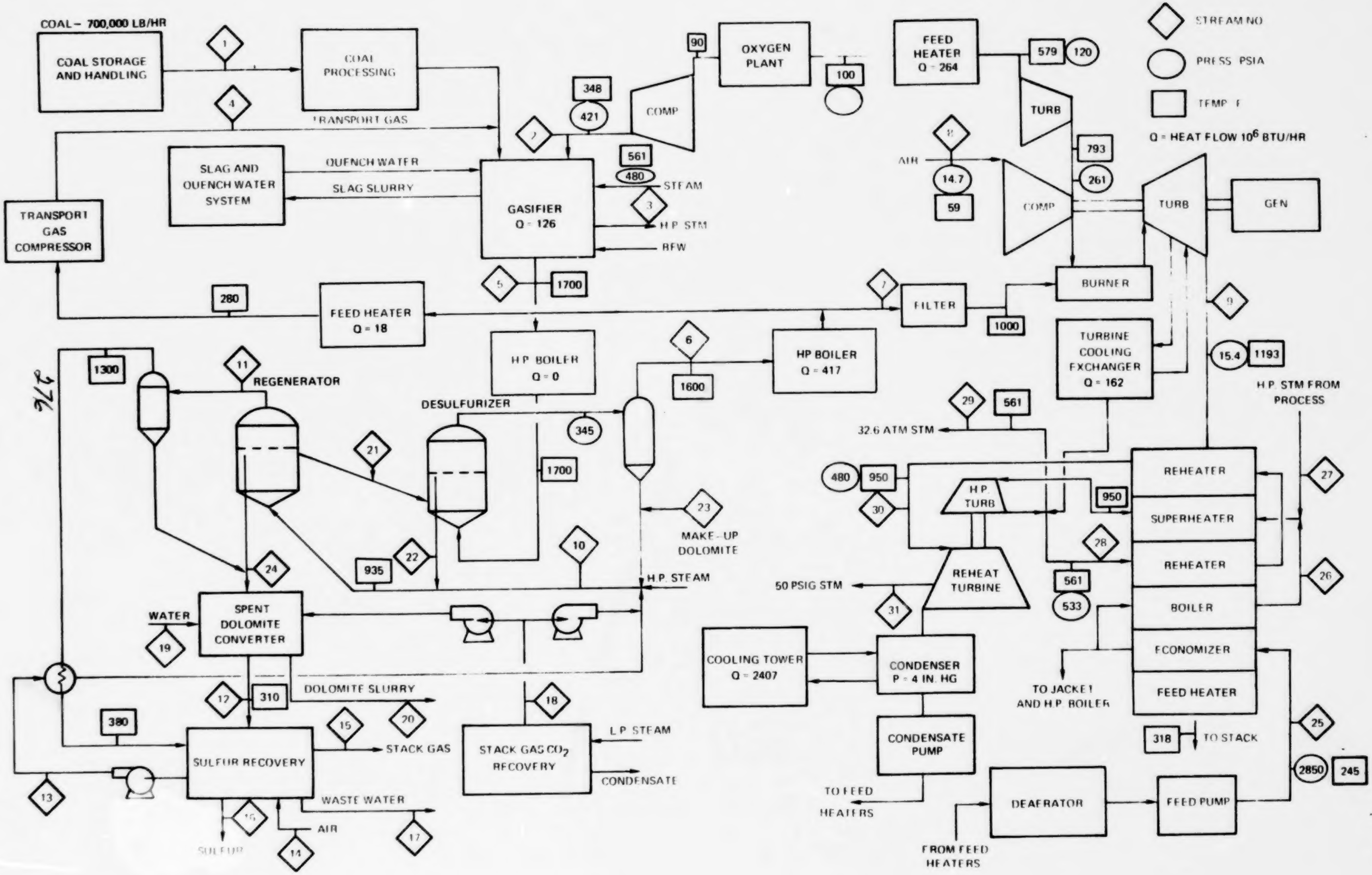
COAL– 700,000 LB/HR
COAL STORAGE AND HANDLING
1
COAL PROCESSING
4
TRANSPORT GAS
TRANSPORT GAS COMPRESSOR
SLAG AND QUENCH WATER SYSTEM
QUENCH WATER
SLAG SLURRY
2
348
421
COMP
90
OXYGEN PLANT
100
561
480
STEAM
3
H.P. STM
BFW
GASIFIER Q = 126
5
1700
280
FEED HEATER Q = 18
H.P. BOILER Q = 0
6
1600
HP BOILER Q = 417
7
FILTER
1000
BURNER
FEED HEATER Q = 264
579
120
TURB
793
8
AIR
14.7
59
COMP
261
TURB
GEN
STREAM NO
PRESS PSIA
TEMP F
Q = HEAT FLOW 10⁶ BTU/HR
9
15.4
1193
TURBINE COOLING EXCHANGER Q = 162
H.P. STM FROM PROCESS
1300
11
REGENERATOR
DESULFURIZER
345
21
1700
23
MAKE-UP DOLOMITE
22
24
935
10
H.P. STEAM
WATER
19
SPENT DOLOMITE CONVERTER
12
310
DOLOMITE SLURRY
20
18
380
15
STACK GAS
SULFUR RECOVERY
STACK GAS CO₂ RECOVERY
L.P. STEAM
CONDENSATE
13
16
SULFUR
14
AIR
WASTE WATER
17
29
561
32.6 ATM STM
480
950
30
H.P. TURB
950
28
561
533
REHEAT TURBINE
50 PSIG STM
31
COOLING TOWER Q = 2407
CONDENSER P = 4 IN. HG
CONDENSATE PUMP
TO FEED HEATERS
DEAERATOR
FROM FEED HEATERS
FEED PUMP
REHEATER
SUPERHEATER
REHEATER
BOILER
ECONOMIZER
FEED HEATER
27
26
25
TO JACKET AND H.P. BOILER
318
TO STACK
2850
245

FIG.11–8

TABLE 11-15

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/CONOCO

Stream		1		2		3		4	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04							2,876	179.3
H_2	2.016							1,810	897.7
CO	28.01							28,772	1,027.2
CO_2	44.01							18,726	425.5
H_2S	34.08							116	3.4
COS	60.07							30	0.5
NH_3	17.03							339	19.9
N_2	28.02			7,302	260.6			373	13.3
O_2	32.00			408,627	12,769.6				
H_2O	18.02					418,320	23,214.2	7,805	433.1
Coal Char		700,000							
TOTAL		700,000		415,929	13,030.2	418,320	23,214.2	60,847	2,999.9

TABLE 11-15 (Cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/CONOCO

Stream		5		6		7		8	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04	72,531	4,521.9	72,531	4,521.9	69,655	4,342.6		
H_2	2.016	44,549	22,097.6	45,652	22,645.0	43,843	21,747.3		
CO	28.01	741,108	26,458.7	725,776	25,911.3	697,004	24,884.1		
CO_2	44.01	412,356	9,369.6	472,373	10,733.3	453,646	10,307.8		
H_2S	34.08	26,255	770.4	2,887	84.7	2,771	81.3		
COS	60.07	4,620	76.9	697	11.6	667	11.1		
NH_3	17.03	8,561	502.7	8,561	502.7	8,222	482.8		
N_2	28.02	9,426	336.4	9,476	336.4	9,053	323.1	11,692,634	417,296
H_2O	18.02	194,380	10,786.9	196,872	10,925.2	189,068	10,492.1		
O_2	32.00							3,547,520	110,860
TOTAL		1,513,786	74,921.1	1,534,775	75,672.1	1,473,929	72,672.2	15,240,154	528,156

Stream		9		10		11		12	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01	1,740,860	39,556	377,351	8,574.2	354,391	8,052.5	2,531	57.5
H_2S	34.08			2,842	83.4	22,012	733.5	3,439	100.9
COS	60.07								
NH_3	17.03								
N_2	28.02	10,134,049	361,672						
H_2O	18.02	737,883	40,948	84,267	4,676.3	725,552	4,026.2	342	19.0
O_2	32.00	2,045,408	63,919						
TOTAL		14,658,200	506,095	464,460	13,333.9	1,101,955	12,812.2	6,312	177.4

TABLE 11-15 (Cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/CONOCO

Stream		13		14		15		16	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
SO_2	64.06					295	4.6		
CH_4	16.04								
S	32.06							23,930	746.4
H_2	2.016								
CO	28.01								
CO_2	44.01	356,921	8,110.0						
H_2S	34.08	2,842	83.4						
COS	60.07								
NH_3	17.03								
N_2	28.02			44,073	1,572.9	44,073	1,572.9		
H_2O	18.02	51,606	2,863.8			67	3.7		
O_2	32.00			13,379	418.1	1,216	38.0		
TOTAL		411,369	11,057.2	57,452	1,991.0	45,651	1,619.2	23,930	746.4

Stream		17		18		19			
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
CH_4	16.04								
H_2	2.016								
CO	28.01								
CO_2	44.01			33,052	751.0				
H_2S	34.08								
COS	60.07								
NH_3	17.03								
N_2	28.02								
H_2O	18.02	34,755	1,928.7	2,483	137.8	32,762	1,818.1		
S	32.06								
TOTAL		34,755	1,928.7	35,535	888.8	32,762	1,818.1		

TABLE 11-15 (cont'd)

MATERIAL BALANCE FOR BCR/OXYGEN BLOWN/CONOCO

Stream		20		21		22		23	
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
	184.01	23,627	128.4					23,627	128.4
	112.46			567,383	5,045.2	651,841	5,796.2		
	140.41			192,867	1,373.6	87,419	622.6		
Inert Solids	100.0	2,180	21.8	109,120	1,091.2	109,120	1,091.2	2,180	21.8
H_2O	18.02	31,549	1,750.8						
TOTAL		57,356	1,901.0	869,370	7,510.0	848,380	7,510.0	25,807	150.2

Stream		24							
Comp	M.W.	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR	LB/HR	MOLS/HR
	112.46	11,347	100.9						
	140.41	3,861	27.5						
Inert Solids	100.0	2,180	21.8						
TOTAL		17,388	150.2						

TABLE 11-16

UTILITY SUMMARY - BCR/CONOCO - AIR-BLOWN

	Coal Prep.	Gasifier	Acid Gas Removal	Sulfur Recovery	CO_2 Supply	Sour Water Stripper	Total
Steam - #/hr							
450 psig (Sat)		120,960	35,425				156,385
100 psig (Sat)					63,455	8,400	71,855
Power - kW	4,360	1,960	470	4,260	9,975	186	21,211
Cooling Water-GPM		5,500	1,225	6,720	5,800	65	19,310
Cooling Duty MM Btu/hr		55.0	12.3	67.2	58.0	0.7	193
Export Heat							
MM Btu			62.1	49.8			--
Temp - F			300	350			--

TABLE 11-17

UTILITY SUMMARY - BCR/CONOCO - O_2 - BLOWN

	Coal Prep.	Gasifier	Acid Gas Removal	Sulfur Recovery	CO_2 Supply	Sour Water Stripper	Total
Steam - #/hr							
450 psig (SAT)		418,320	31,740				450,060
100 psig (SAT)					58,505	7,900	66,400
Power - KW	4,360	1,260	435	3,930	9,200	175	19,360
Cooling Water-GPM		5,500	1,130	6,200	5,350	60	18,240
Cooling Duty-MM Btu/hr		55	11.3	62.	53.5	.6	182
Export Heat -							
MM Btu			57.2	45.9			
Temp - F			300	350			

TABLE 11-18

UTILITY SUMMARY - BCR/CONOCO - AIR-BLOWN

	Coal Prep.	Gasifier	Acid Gas Removal	Sulfur Recovery	CO_2 Supply	Sour Water Stripper	Total
Steam - #/hr 450 psig (SAT)		137,795	35,050				172,845
100 psig (SAT)					62,760	8,400	71,160
Power - kW	2,500	810	465	4,220	9,865	185	18,045
Cooling Water - GPM			1,210	6,650	5,750	65	13,675
Cooling Duty - MM Btu/hr			12.1	66.5	57.5	0.6	137
Export Heat - MM Btu/hr			61.4	49.2			
Temperature - F			300	350			

NOTE: Based on 700,000 #/hr coal feed rate.

a heat recovery boiler is used to drop the raw gas temperature by 100 F. The net effect of the reaction in the absorber is endothermic. In addition, heat leaks and the effect of the recirculated stone combine to cause an approximately 100 F drop in the gas temperature from inlet to outlet of the desulfurizer.

Following desulfurization, gas temperature is dropped to 1000 F by raising high-pressure steam in a waste heat boiler. The steam cycle arrangement is the same as in the low-temperature systems; however, utility requirements of the Conoco system differ somewhat.

High-pressure steam is needed to rehumdify the gas stream to the acceptor regenerator. Low-pressure steam is needed in the CO_2 supply system which removes CO_2 from the stack gas. The other need for low-pressure steam is in the sour water stripper. Overall, the process exports heat to the power system. Heat is available from the acid gas stream that must be cooled before being fed to the liquid phase Claus plant. Also, the waste heat from the sulfur combustor is available.

In the earlier studies, CO_2 for use in acceptor regeneration and conversion of the spent dolomite was removed at pressure from the cool transport gas. In the present study, transport gas flow rate and CO_2 concentration are so low as to preclude use of that stream for a CO_2 supply; therefore, a stack gas bleed stream was selected for use. Less than 3 percent of the stack gas is required so that the resultant change in mixed mean temperature is insignificant when the bleed stream is returned to the stack.

Performance of the high-temperature systems, as expected, showed a somewhat higher overall efficiency than the low-temperature counterparts. However, the advantage was markedly reduced over previous studies where as much as a six point differential had been noted. The comparison between air- and oxygen-blown operation seems to be relatively unaffected by the cleanup system. A slight advantage for the oxygen-blown gasifier with Conoco cleanup was noted due apparently to the increased water vapor in the raw gas from the oxygen-blown gasifier.

The reduced differential between high- and low-temperature desulfurization is of special interest because it results from operation of the gasifier at conditions favorable to combined-cycle performance. Operation under low-steam feed rates has yet to be demonstrated and there is some concern over the ability to control reactor conditions without large quantities of excess steam. Of importance, however, is the fact that there is more than one way to achieve improved overall power plant generating efficiency.

SYSTEM COST ESTIMATES

The following paragraphs discuss the costs of the eight integrated systems and the means by which they were determined. All capital costs are in mid-1977 dollars. Where necessary, costs have been escalated to that time by established procedures (References 11-2, 11-3, and 11-4). In determining the cost of electricity, a relatively simple procedure was used. Yearly capital charges are taken to be 17 percent of the total investment cost. A load factor of 70 percent was assumed and coal cost taken to be $\$1.00/10^6$ Btu. Based upon prior studies, yearly maintenance and operating costs were estimated to be 3.5 percent of the combined-cycle power system cost and 8.5 percent of the fuel processing system cost.

Sources of cost data are discussed below but, in general, gasifier and fuel processing costs for systems with Selexol cleanup were prepared by Fluor Engineers and Constructors; Selexol costs by Allied Chemical; and power systems costs by United Technologies Research Center. Costs for the molten salt system were developed previously at Pullman Kellogg (Reference 11-5) and for the Conoco desulfurization process by Conoco (Reference 11-6) under contract to the EPA.

A summary of fuel processing system costs is given in Table 11-19 for low-temperature and in Table 11-20 for high-temperature cleanup systems. Power system costs are given in Table 11-21. While total costs are on the same basis, individual costs for the various parts of the fuel systems include indirect costs for engineering and supervision during construction but do not include any funds for contingency. For the power systems, indirect costs are not included in the individual items.

Contingencies have been applied to each group of costs. These vary based on judgment of the overall development status of the equipment. For the power systems a contingency of 15 percent was used while the factor was higher for the fuel systems, 21 percent for low- and 25 percent for high-temperature systems.

Fuel Processing Systems with Selexol Desulfurization

For this group of systems, cost estimates were prepared by Fluor. They are based on data developed previously under contract to EPRI and adjusted to the parameters of this study. An example of the costing basis is given in Appendix B for the air-blown BCR system. The data provided by Fluor covered all major fuel processing equipment with the exception of the Selexol system.

The cost of that unit was estimated by the developer, Allied Chemical, and scaled as a function of gas flow rates. The costs appear to be consistent with data that have subsequently been obtained from Fluor. Other costs assigned to the fuel processing section include the bleed-air cooling heat exchanger and boost compressor, an incremental cost for cooling water (cooling tower plus circulating system) and boiler make-up water treatment facilities (for treating gasifier steam), and "other support facilities". This last category includes waste water treatment, gas flaring, instrument air, etc. For the oxygen-blown system, an auxiliary compressor was included for startup of the air separation plant.

Molten Salt System Costs

For the molten salt system, cost information was provided by Pullman Kellogg (Reference 11-5). The estimate is considered to be a rough order of magnitude number and required escalation and scaling to be comparable with

TABLE 11-19

FUEL PROCESSING COST SUMMARY
LOW-TEMPERATURE CLEANUP

	U-Gas EPRI Data	U-Gas Low Stm	BCR Air	BCR O_2	Molten Salt
Coal Handling	16,725	16,725	28,287	28,287	*137,900
Air Separation				41,155	
Oxidant Feed & Cooling	4,241	4,042	3,861	7,622	
Gasification & Ash Handling	30,699	30,699	40,353	35,040	
Gas Cooling & Scrub	53,005	53,005	31,040	19,605	
Acid Gas Removal	29,000	27,500	26,600	16,600	
Sulfur Rec. & Tail Gas Cleanup	10,894	10,894	10,894	10,894	10,260
Sour Water Treating	1,853	1,853	7,377	7,377	-
Cooling & Make-up Water	12,064	5,547	4,661	11,027	2,468
Other Support Facilities	8,400	8,400	10,400	10,400	8,400
Aux. O_2 Plant Compressor				2,245	
Subtotal	166,881	158,665	163,473	188,752	159,028
Contingency	35,045	33,320	34,329	39,638	33,396
Escalation During Constr.	27,986	26,608	27,414	31,654	26,717
Subtotal	229,912	218,593	225,216	260,044	219,141
Interest During Constr.	53,340	50,714	52,250	60,330	50,841
Total Cost	283,252	269,307	277,466	320,374	269,982
Costs in Thousands of Dollars - Mid 1977					
Unit Cost - $/kW	264	243	253	307	283

* Total cost for gasifier and cleanup system

TABLE 11-20

FUEL PROCESSING COST SUMMARY
HIGH-TEMPERATURE CLEANUP

	U-Gas Low Stm	BCR-Air	BCR-O_2
Coal Handling	16,725	28,287	28,287
Air Separation			41,155
Oxidant Feed and Cooling	4,042	3,861	7,622
Gasification & Ash Handling	30,699	40,353	35,040
Gas Cooling	9,720	10,160	7,700
Acid Gas Removal	16,444	16,444	16,444
Particulate Removal	8,529	8,313	6,210
Sulfur Recovery	12,279	12,279	12,279
CO_2 Supply	3,676	3,676	3,676
Cooling & Makeup Water	4,449	4,451	10,516
Other Support Facilities	8,400	10,400	10,400
Aux. O_2 Plant Compressor			2,245
Subtotal	114,963	138,224	181,574
Contingency (30%)	34,489	41,467	54,472
Escalation During Constr.(.168)	19,314	23,222	30,504
Subtotal	168,766	202,913	266,550
Interest During Const. (.232)	39,154	47,076	61,840
Total Cost	207,920	249,989	328,390

Costs in thousands of dollars - Mid 1977

Unit Cost - \$/kW	182	220	299

TABLE 11-21

POWER SYSTEMS COST SUMMARY

Item	U-Gas Selexol	U-Gas Selex (Low Stm)	BCR-Air Selexol	BCR-O_2 Selexol	Molten Salt	U-Gas Conoco Low-Stm	BCR-Air Conoco	BCR-O_2 Conoco
341 Site and Structures	10,577	10,771	10,702	10,445	9,487	10,825	10,838	10,620
343 Gas Turbine	57,763	58,874	59,831	55,396	51,571	60,086	61,062	57,698
344 Gas Generator	15,289	15,618	15,903	14,456	13,556	16,126	16,298	15,202
312 Boiler Plant	72,165	75,698	77,609	82,046	63,494	60,062	66,953	84,058
314 Steam Turbogenerator	37,541	38,029	36,876	36,544	37,735	38,043	37,360	36,957
345 & 353 Acc. Elect. Equip.	18,547	19,187	18,853	17,523	13,169	19,463	19,391	18,299
346 Misc. Pwr Plant Equip.	559	568	563	541	528	573	571	554
Adj. Direct Const. Costs	212,441	218,745	220,337	216,951	189,540	205,178	212,473	223,388
Other Expenses	18,589	19,140	19,279	18,983	16,585	17,953	18,591	19,547
Direct Const. Costs	231,030	237,885	239,616	235,934	206,125	223,131	231,064	242,935
Indirects	80,861	83,260	83,866	82,577	72,144	78,096	80,872	85,027
Escalation	35,690	36,749	37,017	36,448	31,464	34,470	35,696	37,529
Total Investment Cost	347,581	357,894	360,499	354,959	309,733	335,697	347,632	365,491
Interest During Const.	80,639	83,031	83,636	82,350	71,858	77,882	80,651	84,794
Grand Total	428,220	440,925	444,135	437,309	381,591	413,579	428,283	450,285

Costs are in thousands of dollars - Mid 1977

Unit cost - $/kW	399	398	404	418	400	361	377	410

other data. Therefore, the degree of confidence in the data is low. However, the cost in $/kW is quite consistent with the other low-temperature systems.

Those items not covered in the estimate provided by Pullman Kellogg were scaled on a consistent basis with the other low-temperature systems.

High Temperature Fuel Processing System Costs

For these systems, coal handling and gasification costs are the same as for the low-temperature systems. Conoco costs were obtained from data presented in Reference 11-6. However, because gas cooling requirements differ significantly, that equipment was removed from the Conoco process and treated separately. Costs were assumed to vary with the 0.7 power for sulfur removal, sulfur recovery and make-up CO_2 sections. Data contained in Reference 11-7 by Stone and Webster were helpful in estimating costs of hot gas handling equipment, i.e., the high-pressure boiler and particulate filter. Information contained there correlated well with other sources (References 11-8 and 11-9) and was used for those estimates.

While the degree of confidence in the costs of the three high-temperature systems is not high, they are valuable for comparative purposes. The U-Gas process shows a cost advantage in both coal preparation and gasification over the BCR air-blown system. These two areas account for most of the difference between the two. In comparing the air- and oxygen-blown BCR systems, all costs are similar with the exception of the air-separation plant. In the low-temperature comparison, the cost of air-separation was offset in large measure by reduced gas cooling and desulfurization costs. Here, desulfurization costs were based directly on sulfur removal rate as the desulfurization process is not aided by increased partial pressure of H_2S and the major part of the Conoco system is devoted to regeneration of the dolomite. Thus, the capital cost and overall generating cost for the high-temperature oxygen-blown system does not show any significant advantage over the low-temperature system.

Power System Costs

Power system costs are relatively constant on a per kilowatt basis. They are summarized in Table 11-21. Using the air-blown BCR Selexol system as an example, the costing breakdown is presented in Appendix C. As mentioned in the discussion of performance, the use of a fixed steam cycle (i.e., a steam cycle whose operating conditions have not been optimized for this particular application) tends to increase the size of the gas turbine exhaust waste heat boiler in the oxygen-blown systems. Also, because it was necessary to use the gas turbine exhaust waste heat boiler for feedwater heating in the two air-blown Conoco systems, temperature differences were increased and boiler costs reduced significantly. This suggests that some further optimization is possible, but it is doubtful that the resulting improvement would materially change the study's conclusions.

The power systems presented in this report differ significantly as compared to those of earlier studies (References 11-1 and 11-10). The changes are due to the decision to use a moderate (18:1) pressure ratio gas turbine so that a 2400 psi reheat steam cycle could be utilized. This results in improved cycle efficiency and an overall reduction in generating costs, despite the fact that the steam generation fraction is increased. This has been discussed previously in Section 7. Turbogenerator costs, as compared to Reference 11-10, increase but not out of proportion to the increased output. In fact, the unit price is slightly less after taking into account escalation and the reduction in cost due to increased size. However, waste heat boiler sizes are up by 50 to 60 percent and when combined with revised unit cost data that show a higher than average escalation rate, waste heat boiler costs have increased approximately by a factor of two. The overall effect is a slight increase in unit cost of the equipment associated with steam power generation.

Unfortunately, the picture is clouded by the fact that some of the steam associated costs are assigned to the gasification and cleanup portion of the system. These are primarily the costs of steam generation in the gasifier, the fuel gas heat recovery boiler, and boost compressor cooling boiler. If these were included, the unit cost of the steam power generation would probably decrease, or at worst remain constant, because of the improved efficiency achieved with a reheat cycle.

Gas turbine manufacturing costs appear to follow overall escalation factors quite closely. Other considerations, however, indicate that current mark-up factors may be insufficient when applied to advanced, high-temperature, high-performance engines. These improvements generally tend to reduce or at least not increase the unit manufacturing cost. On the other hand the development costs needed to produce the improvements are sharply increased. The number of units over which these costs can be spread is one of the major uncertainties. Also, the availability of government funding for development of these improvements is questionable. Because of these uncertainties, recent studies have indicated that estimates of gas turbine prices (based on the availability of advanced technology) should be increased by some 70 percent over those used in previous studies although the inflation rate for this area has been approximately 15 percent.

Current conditions are such that it is prudent to raise the total of indirect costs from 23 to 30 percent. These consist of a contingency allowance plus engineering and construction supervision costs. The overall result is a relatively large increase in the unit cost of the power system. However, it is balanced by an improved system efficiency and the resulting higher output tends to reduce the unit cost of fuel processing.

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APPENDIX A

WATER TREATMENT AND REUSE

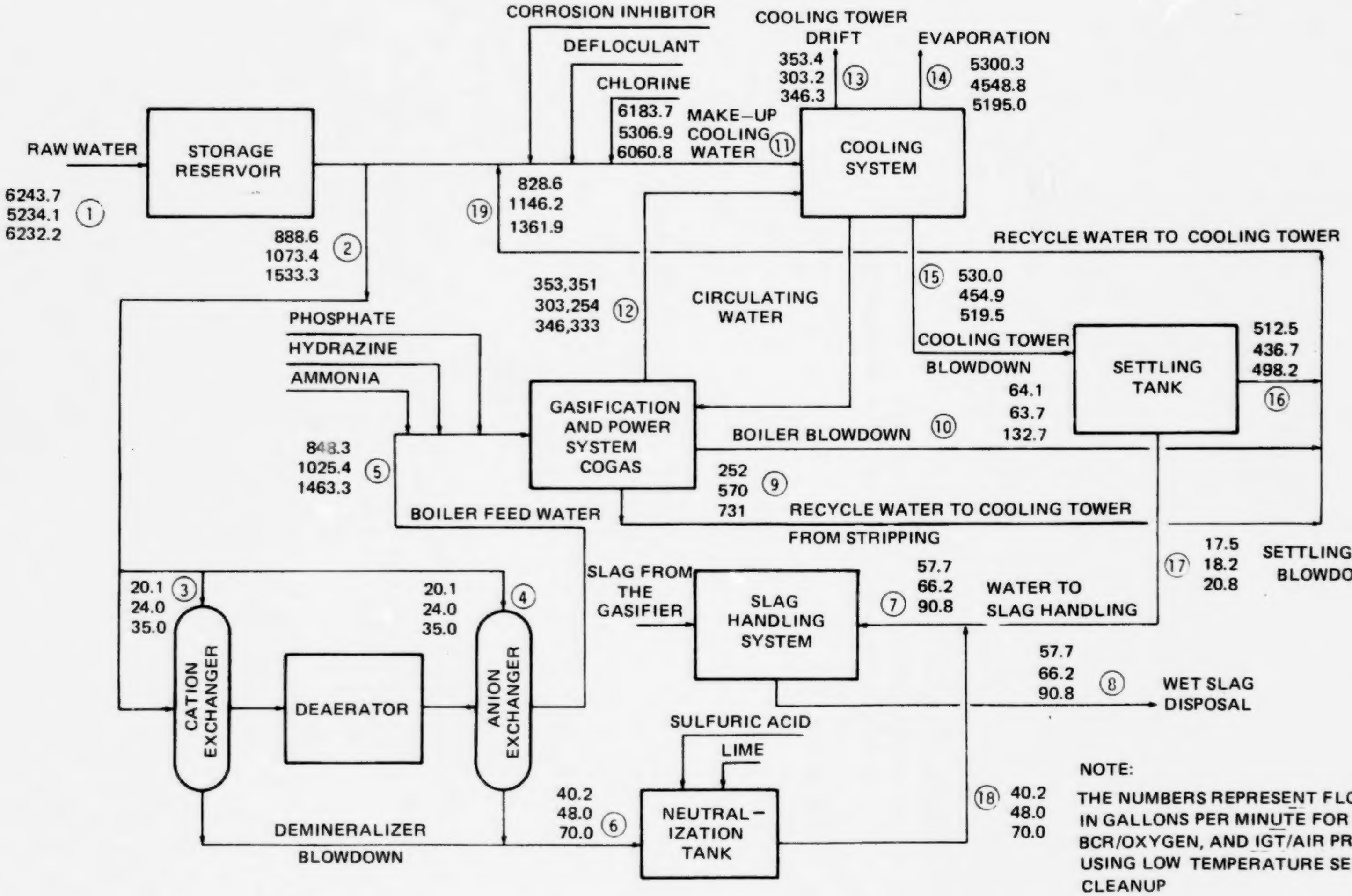
The following appendix presents a discussion of the water requirements, treatment and reuse for the integrated power plants described in the current study.

WATER REUSE FOR COGAS SYSTEMS

High water requirements for coal conversion plants impose certain constraints on water utilization. The water use and reuse system should be designed to make maximum utilization of scarce water without damaging the environment. The water for the plant will be provided from the reservoir, which will be withdrawn from a local source. The water supply system will be capable of delivering sufficient water to match the plant requirements. The commercial plant will probably have to meet zero discharge criteria by the year 1985. None of the water will then be returned to any stream or stream bed. The COGAS plant water reuse system is shown in Figure A-1 for the BCR/Air-Blown/Selexol, BCR/Oxygen-Blown/Selexol, and U-Gas/Air-Blown/Selexol processes. The flowsheet is greatly simplified and is intended to show the overall plant water balance or overall recirculating systems. The water balances for the above three processes are given in Tables A-1 to 3. The numbers represent combined flow rates of cooling water for condenser and gasification process units. Cooling water requirements for the gasification process units are given separately at the bottom of each table. The cooling water requirements for BCR and IGT processes employing Conoco cleanup are shown in Tables A-4 to 6. The cooling water requirements for the gasification units, acid gas removal, sulfur recovery, carbon dioxide supply, and sour water stripper are shown at the bottom of each table.

The processes will use two separate cooling tower systems. The first system of cooling towers will supply water to the surface condenser. The second system will supply cooling water to the gasification system. The make-up water requirements for the BCR/Air-Blown/Conoco, BCR/Oxygen-Blown/Conoco and IGT/Air-Blown/Conoco combined cooling systems are about 6,460, 6,150, and

COMBINED CYCLE–POWER PLANT – WATER TREATMENT AND REUSE


RAW WATER
6243.7
5234.1
6232.2
①
STORAGE RESERVOIR
888.6
1073.4
1533.3
②
CORROSION INHIBITOR
DEFLOCULANT
CHLORINE
6183.7
5306.9
6060.8
MAKE–UP COOLING WATER
⑪
828.6
1146.2
1361.9
⑲
COOLING TOWER DRIFT
353.4
303.2
346.3
⑬
EVAPORATION
⑭
5300.3
4548.8
5195.0
COOLING SYSTEM
RECYCLE WATER TO COOLING TOWER
⑮
530.0
454.9
519.5
COOLING TOWER BLOWDOWN
64.1
63.7
132.7
SETTLING TANK
512.5
436.7
498.2
⑯
353,351
303,254
346,333
⑫
CIRCULATING WATER
PHOSPHATE
HYDRAZINE
AMMONIA
GASIFICATION AND POWER SYSTEM COGAS
BOILER BLOWDOWN
⑩
848.3
1025.4
1463.3
⑤
252
570
731
⑨
RECYCLE WATER TO COOLING TOWER
FROM STRIPPING
BOILER FEED WATER
17.5
18.2
20.8
⑰
SETTLING TANK BLOWDOWN
SLAG FROM THE GASIFIER
SLAG HANDLING SYSTEM
57.7
66.2
90.8
⑦
WATER TO SLAG HANDLING
57.7
66.2
90.8
⑧
WET SLAG DISPOSAL
20.1
24.0
35.0
③
CATION EXCHANGER
DEAERATOR
20.1
24.0
35.0
④
ANION EXCHANGER
SULFURIC ACID
LIME
40.2
48.0
70.0
⑥
NEUTRAL–IZATION TANK
⑱
40.2
48.0
70.0
DEMINERALIZER BLOWDOWN

NOTE:

THE NUMBERS REPRESENT FLOWRATES IN GALLONS PER MINUTE FOR BCR/AIR, BCR/OXYGEN, AND IGT/AIR PROCESSES USING LOW TEMPERATURE SELEXOL CLEANUP

FIG. A–1

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TABLE A-1

WATER BALANCES FOR THE BCR/AIR-BLOWN/SELEXOL PROCESS OVERALL WATER BALANCE (SEE FIG. A-1)

Stream No.	Description	Flows, lb/hr
1	Raw water	3,120,600
2	Raw water to demineralizer	444,120
3	Acid regenerant and rinse water	10,050
4	Alkaline regenerant and rinse water	10,050
5	Boiler feed water to process	423,980
6	Demineralizer blowdown	20,090
7	Water to slag handling	28,840
8	Water associated with slag disposal	28,840
9	Make-up water to cooling tower from stripping	125,950
10	Boiler blowdown	32,040
11	Make-up cooling water	3,090,610
12	Cooling tower circulating water	176,604,880
13	Drift losses from cooling water	176,630
14	Evaporation losses from cooling tower	2,649,090
15	Cooling tower blowdown	264,890
16	Make-up water to cooling tower from settling tank	256,150
17	Settling tank blowdown to slag handling	8,750
18	Water from neutralization tank to slag handling	20,100
19	Recycle water to cooling tower as make-up water	414,130

Cooling Water Requirements for Process Units

1.	Gasification	2,739,400
2.	Gas scrubbing	1,265,990
3.	Acid gas removal	13,879,450
4.	Ammonia recovery	323,870
5.	Sulfur recovery	62,980

TABLE A-2

WATER BALANCES FOR THE BCR/OXYGEN-BLOWN/SELEXOL PROCESS OVERALL WATER BALANCE (SEE FIG. A-1)

Stream No.	Description	Flow lb/hr
1	Raw water	2,616,000
2	Raw water to demineralizer	536,490
3	Acid regenerant and rinse water	12,000
4	Alkaline regenerant and rinse water	512,600
5	Boiler feed water to process	23,990
6	Demineralizer blowdown	83,090
7	Water to slag handling	83,090
8	Water associated with slag disposal	284,890
9	Make-up water to cooling tower from stripping	31,840
10	Boiler blowdown	2,652,410
11	Make-up cooling water	151,566,350
12	Cooling tower circulating water	151,540
13	Drift losses from cooling tower	227,360
14	Evaporation losses from cooling tower	2,273,490
15	Cooling tower blowdown	218,260
16	Make-up water to cooling tower from settling tank	211,650
17	Settling tank blowdown to slag handling	9,100
18	Water from neutralization tank to slag handling	23,990
19	Recycle water to cooling tower as make-up water	572,870

Cooling Water Requirements for Process Units

1.	Acid gas removal	7,892,840
2.	Ammonia recovery	143,940
3.	Sulfur recovery	62,980

TABLE A-3

WATER BALANCES FOR THE IGT/AIR-BLOWN/SELEXOL PROCESS OVERALL WATER BALANCE (SEE FIG. A-1)

Stream No.	Description	Flow lb/hr
1	Raw water	3,114,850
2	Raw water to demineralizer	766,340
3	Acid regenerant and rinse water	17,490
4	Alkaline regenerant and rinse water	17,490
5	Boiler feed water to process	731,360
6	Demineralizer blowdown	34,990
7	Water to slag handling	45,380
8	Water associated with slag disposal (54% solids)	45,380
9	Make-up water to cooling tower from stripping	365,350
10	Boiler blowdown	66,320
11	Make-up cooling water	3,029,200
12	Cooling tower circulating water	173,097,230
13	Drift losses from cooling tower	173,080
14	Evaporation losses from cooling tower	2,596,460
15	Cooling tower blowdown	259,660
16	Make-up water to cooling tower from settling tank	249,000
17	Settling tank blowdown to slag handling	10,400
18	Water from neutralization tank to slag handling	34,990
19	Recycle water to cooling tower as make-up water	680,680

Cooling Water Requirements for Process Units

1.	Gas scrubbing	1,780,290
2.	Acid gas removal	151,139,950
3.	Ammonia recovery	402,840

TABLE A-4

COOLING WATER REQUIREMENTS FOR BCR/AIR-BLOWN/CONOCO PROCESS

Description	Flow, lb/hr
Boiler feed waters to demineralizers	242,820
Make-up cooling water	3,229,300
Cooling tower circulating water	184,531,140
Evaporation losses from cooling tower	2,767,970
Cooling tower blowdown	276,800
Drift losses from cooling tower	184,530

Cooling Water Requirements for process Gasification Units

Acid gas removal	612,260
Sulfur removal	3,358,660
CO_2 supply	2,898,840
Sour water stripper	32,490
Gasifier	2,748,900

TABLE A-5

COOLING WATER REQUIREMENTS FOR BCR/OXYGEN-BLOWN/CONOCO PROCESS

Description	Flows, lb/hr
Boiler feed water to demineralizers	570,010
Make-up cooling water	3,073,130
Cooling tower circulating water	175,607,450
Evaporation losses from cooling tower	2,634,110
Cooling tower blowdown	263,410
Drift losses from cooling tower	175,610

Cooling Water Requirements for Process Gasification Units

Acid gas removal	564,770
Sulfur recovery	3,098,760
CO_2 supply	2,673,930
Sour water stripper	29,990

TABLE A-6

COOLING WATER REQUIREMENTS FOR IGT/AIR-BLOWN/CONOCO PROCESS

Description	Flows, lb/hr
Boiler feed water to demineralizers	289,093
Make-up cooling water	3,336,457
Cooling tower circulating water	190,654,765
Evaporation losses from cooling tower	2,859,821
Cooling tower blowdown	285,982
Drift losses from cooling tower	190,654

Cooling Water Requirements for Process Gasification Units

Acid gas removal	604,758
Sulfur recovery	3,323,670
CO_2 supply	2,873,850
Sour water stripper	32,487

6,675 gpm, respectively. The make-up water to the cooling tower for gasification systems for the above three processes is about 340, 220, and 240 gpm, respectively. The total cooling water supply (recirculating water) for the gasification systems for each process is BCR/Air-Blown/Conoco, 13,675; BCR/Oxygen-Blown/Conoco, 12,740; and IGT/Air-Blown/Conoco, 19,310 gpm.

The molten salt process, which is different from the above processes, requires a cooling water supply to the power system condenser of about 243,750 gpm. A separate cooling tower supplies 45,700 gpm of cooling water to the gasification system and off-site facilities.

The BCR/Air-Blown/Selexol process and the water system associated with it are described below. The other two processes, BCR/Oxygen-Blown/Selexol and IGT/Air-Blown/Selexol, operate similarly with the exception of differing flow-rate values. Water will enter the raw storage pond at 7,810 gpm. The lined reservoir will hold 170 million gallons, about a two-week holdup. The process will require 6,240 gpm, and approximately 20 percent of the total inflow to raw storage pond, 1,250 gpm, will consist of auxiliary water requirements such as mine and offsite needs, and storage pond evaporation losses.

Boiler Feedwater

The process water from the raw storage ponds is broken into two major streams. One of the streams, having a flowrate of about 889 gpm, flows to the demineralizer; the second goes to the cooling towers. The demineralized water is used for generating the necessary steam for the process. Steam is utilized in the gasifier to supply hydrogen and is used in the ammonia recovery unit for the tower reboilers. Sour water strippers require a steam flow of about 1.0 to 1.2 lb per gallon of sour water entering the unit. About 4.5 percent of the water entering the demineralizers is used as regenerant and rinse water. A high steam flow rate is required to drive the steam turbines of the COGAS system. However, most of the steam is condensed and recycled to the boilers. Only 64 gpm of fresh water will be needed to make-up the boiler blowdown. Modern, high-pressure boilers have blowdown rates of 0.1 to 1.0 percent of the total steam generated.

Cooling Water

Most of the cooling water make-up is the fresh water from the reservoir. Of 6,184 gpm water required for the cooling tower make-up, 5,355 gpm comes from the reservoir. This water will be passed through cold lime treating if the total dissolved solids concentration is high. Since lime treatment for the water is not essential in this study, it is not included in the flowsheets. The rest of the total cooling tower make-up water, about 13 percent, is the water recovered and recycled from the process itself.

Raw gas from the gasifiers will contain unreacted steam, part of which originated from the moisture in the coal and part of which was input for gasificaton. The raw gas is free of any tars and oils, but may contain trace amounts of organics. The gasifier raw gas will be cooled in a heat recovery system and further scrubbed with water to remove ammonia.

The scrubber water will then be stripped to recover ammonia. Part of this treated water is reused in the scrubber and the rest is passed to the cooling tower. Organics that may be present will be degraded in the cooling tower. The water from the stripping section accounts for 4 percent of the total make-up.

The second reuse system in the process consists of the boiler and cooling tower blowdown treatment. Blowdown from all boilers in the plant will be collected and reused as make-up to the cooling tower. All the boilers in the plant will run to a limit of the allowable concentration of silica. Because cooling towers can operate at a higher silica limit than boilers, blowdown from the boilers can be used in the cooling towers. The cooling tower blowdown water will have a high concentration of dissolved and suspended solids. The suspended solids are removed in a settling tank. When the excess water from stripping section and boiler blowdown water is combined with the cooling tower blowdown, the total dissolved solids concentration of the combined recycle water is less than cooling tower blowdown alone, and can be used in the cooling tower without treatment.

Cooling water is required for removing excess heat from process units such as acid gas removal, sulfur recovery and ammonia recovery. A temperature rise of about 15 to 20 F is allowed for the water circulating through the heat exchangers. The heated water is cooled in the cooling towers and then recirculated in the heat exchangers. A separate cooling tower will be used for this purpose.

Slag Handling System

The coal entering the gasifiers will be about 8.7 percent ash. This non-reactive ash will go through the gasifier and leave in the form of slag. The slag, at a high temperature, is cooled by water sprayed at the bottom of the gasifier. The cooled slag is further quenched in a slag quench tank. Rinse water from the demineralizers will be treated in the neutralization tank. This stream and the blowdown from the settling tank can be used to quench the slag. The slag-water mixture can be further concentrated in the clarifier. The water recovered will then be recycled to the quench tank.

SOUR WATER STRIPPING

The raw gas from the gasifier is scrubbed in a water scrubber. The scrubber may be packed-bed tower or a tray tower. Almost all ammonia and part of the H_2S and CO_2 are absorbed in the water. Phenols, cyanides and organics may be present in the raw gas and also will appear in the sour water. Ammonia, a valuable by-product, is recovered by stripping. Since ammonia stripping is an integral part of the gasification process, two commonly used stripping processes, the two-stage All-Distillation Process and the Phosam-W Process, are described below.

Sour Water Strippers

Sour water stripping is used to remove hydrogen sulfide and ammonia from sour waters. In the gasification plants, ammonia is mostly neutralized by carbon dioxide. Ammonia must be removed from the sour water to a level low enough to conform to the effluent regulations for wastewater. It should be collected separately because it interferes with the Claus sulfur recovery process. Since sulfur is a saleable commodity, it should be collected in a salable form. H_2S cannot be vented to the atmosphere; instead, it must be collected and sent to the sulfur recovery plant. Sour water may also contain organic acids, sulfide oxidation products, and phenols and cyanides, some of which interfere with ammonia and hydrogen sulfide separation. Phenols, if present in high concentrations, are removed by solvent extraction prior to stripping. However, phenols are not present in the raw gas of the gasification processes considered in this study. If present, their concentration will be low.

A conventional sour water stipper is shown in Figure A-2. In general, the stripper contains 10 to 20 actual trays. In some cases, packed towers are employed. Heat integration is common between the feed and bottom stream, and the bottoms are either reboiled or stripped with live steam at a rate usually in the range 10-24 lb steam/100 lb water. The stripper overhead is condensed to produce an off-gas suitable for further treatment in a sulfur plant and a reflux which is either returned to the top tray in the tower or the feed drum. Because of its low solubility, hydrogen sulfide can be stripped more readily than ammonia. A temperature of 230°F will remove more than 90 percent of ammonia. However, 90 percent of hydrogen sulfide removal requires a temperature of only 100°F. In a single stage (tower) sour water stripping operation, ammonia must be separated from the CO_2 and H_2S vapors. This may be done by absorbing the ammonia in sulfuric acid, which will produce crystalline ammonium sulfate for sale. Another procedure is to absorb the ammonia in phosphoric acid or ammonium phosphate.

Two-Stage All-Distillation Process

This proprietary process, developed by Chevron Research, uses two stripping columns for the recovery of high-purity ammonia and hydrogen sulfide from sour waters as separate by-products. The process has been in successful commercial use since 1966 and a number of commercial installations are in operation.

The process flow diagram is shown in Figure A-3. The stripped water from the process is 99.9 percent water by weight with less than 50 ppm ammonia and 10 ppm H_2S. The H_2S stream is suitable for feed to plants producing sulfur or sulfuric acid without further treatment. High purity ammonia can be produced in either aqueous or anhydrous liquid form, depending on the desired end-use market.

A typical performance feed rate of 134 gpm is given below (the numbers represent the stream numbers on Figure A-3):

CONVENTIONAL SOUR WATER STRIPPING PROCESS

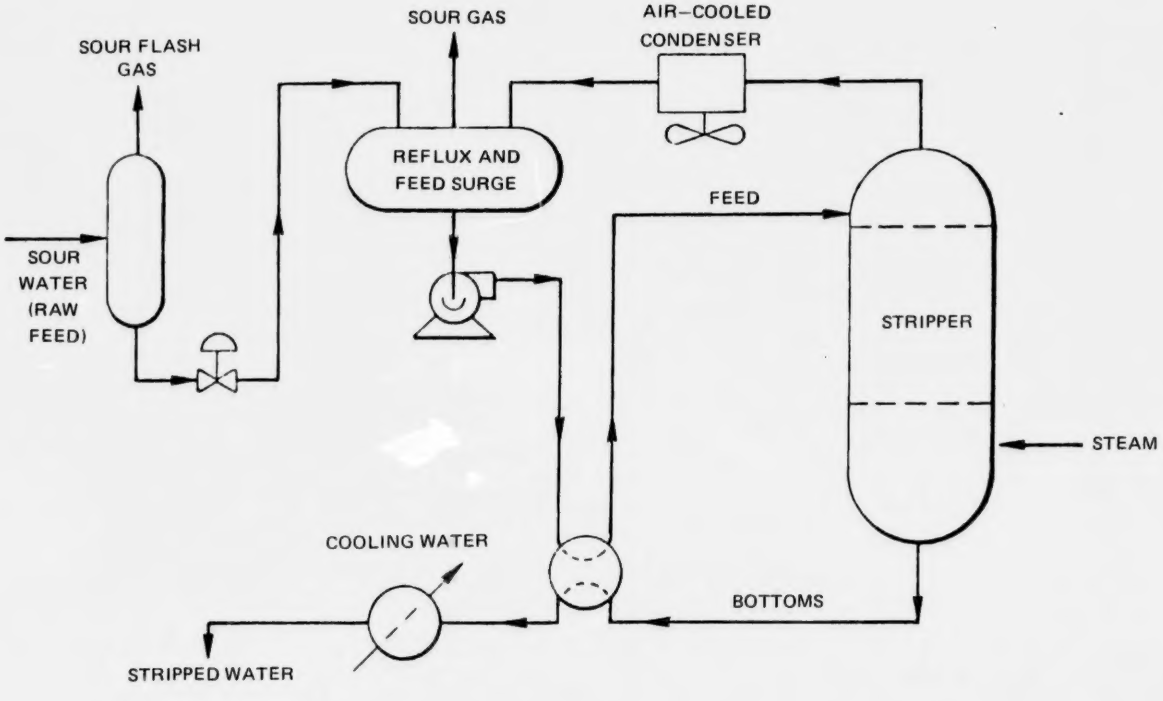
SOUR FLASH GAS
SOUR GAS
AIR-COOLED CONDENSER
REFLUX AND FEED SURGE
FEED
SOUR WATER (RAW FEED)
STRIPPER
STEAM
COOLING WATER
BOTTOMS
STRIPPED WATER

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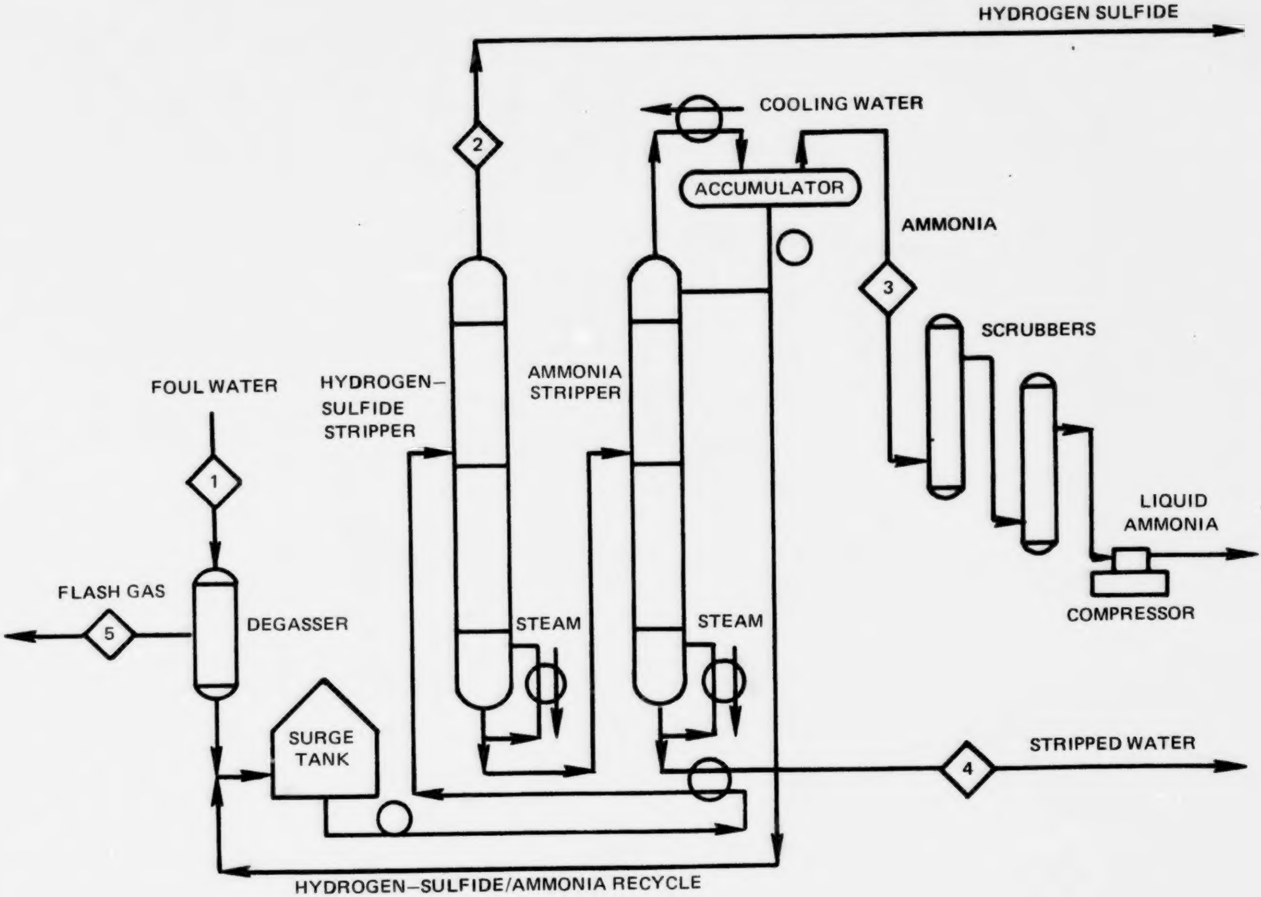
FIG. A–3

TWO STAGE ALL–DISTILLATION PROCESS


HYDROGEN SULFIDE
COOLING WATER
ACCUMULATOR
AMMONIA
SCRUBBERS
FOUL WATER
HYDROGEN–
SULFIDE
STRIPPER
AMMONIA
STRIPPER
LIQUID
AMMONIA
FLASH GAS
DEGASSER
STEAM
STEAM
COMPRESSOR
SURGE
TANK
STRIPPED WATER
HYDROGEN–SULFIDE/AMMONIA RECYCLE
1
2
3
4
5

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Feed (Stream-1)

Feed rate	134 gpm
Dissolved material, by weight	3.8 percent
Dissolved ammonia, by weight	3-5.5 percent
Dissolved H_2S, by weight	4-5.5 percent
Dissolved hydrocarbons	---------------

Stripped H_2S (Stream-2)

Generation rate	20.6 TPD
H_2S by weight	99.9 percent
H_2O by weight	0.1 percent
NH_3 by weight	30 ppm

Stripped NH_3 (Stream-3)

Generation rate	10.3 TPD
NH_3 by weight	99.9 percent
H_2O by weight	0.1 percent
H_2S by weight	1 ppm

Stripped Water (Stream-4)

H_2O by weight	99.9 percent +
H_2S by weight	10 ppm
NH_3 by weight	50 ppm

Flash Gas (Stream-5)

Contains light hydrocarbons like CH_2, C_2H_6

Heat Requirements

3.9×10^5 Btu

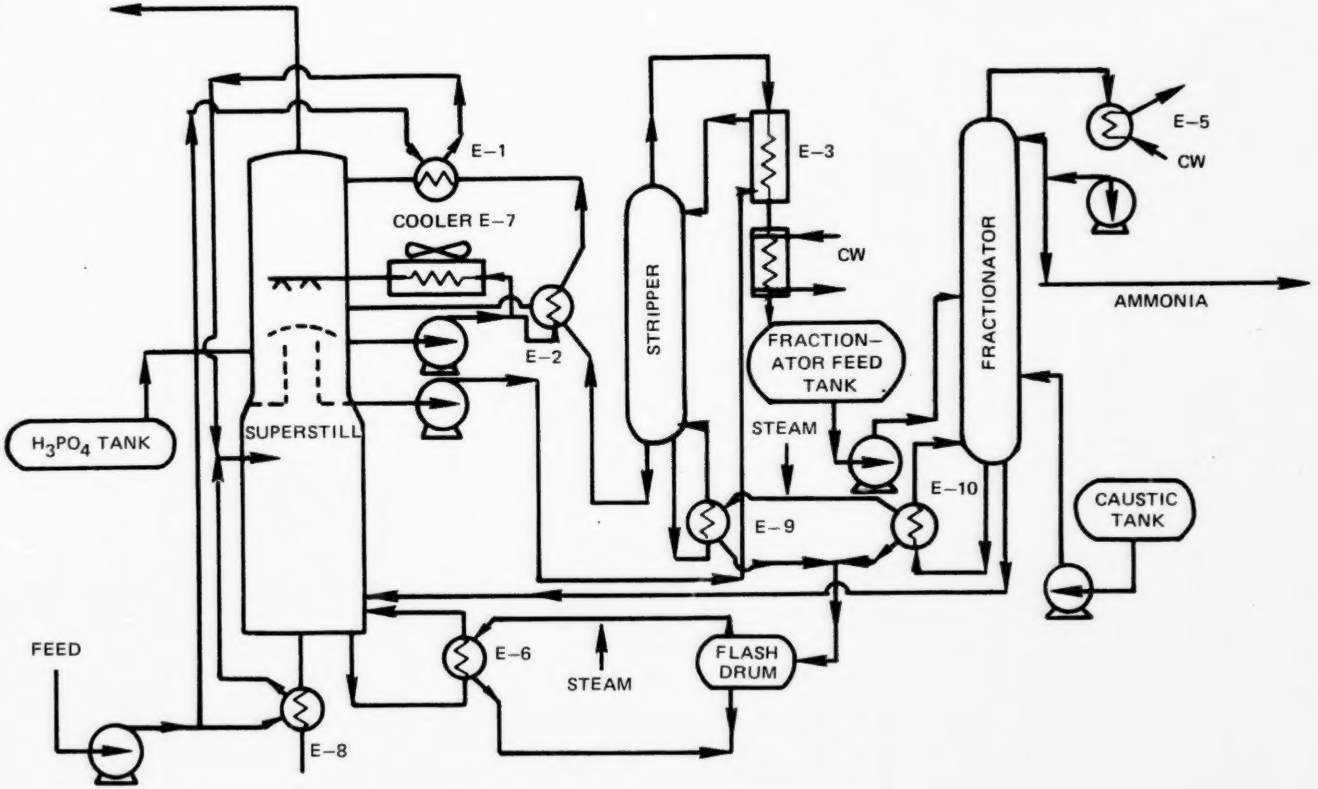
Phosam-W Process

The Phosam-W process, developed by U.S. Steel, removes ammonia from waste-waters containing acid gases by using phosphoric acid as a stripping medium rather than the commonly used sulfuric acid. Phosphoric acid, with three bound hydrogen atoms, can bind ammonium ions tightly enough for absorption, but loosely enough for ammonia recovery via stripping.

Figure A-4 is a flowsheet of the Phosam-W process. Feed, consisting of water with dissolved ammonia and acid gas is preheated in two streams, and recombined to enter the stripping section of the superstill. The low-pressure (4 psig) stripping section removes both ammonia and acid gases from the water. The superstill bottoms, with an ammonia content of less than 200 ppm, is recycled to the preheater to heat part of the wastewater feed. Indirect steam, at 60 psig, provides most of the energy input to the superstill stripper.

FIG. A-4

PHOSAM-W PROCESS FOR AMMONIA SEPARATION


E-1
COOLER E-7
E-2
E-3
CW
E-5
CW
STRIPPER
FRACTIONATOR
AMMONIA
FRACTION-ATOR FEED TANK
STEAM
E-9
E-10
CAUSTIC TANK
H_3PO_4 TANK
SUPERSTILL
FEED
E-8
E-6
STEAM
FLASH DRUM

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Vapor from the superstill stripper section rises, entering the absorber section of the superstill. The vapor is contacted with a spray of cool ammonium phosphate solution. Scrubbing continues in the tray section of the absorber where gas and lean ammonium phosphate solution contact countercurrently.

A solution exchanger in the bottom of the absorber purges acid gases from the rich solution. The rich solution is then pumped to the stripper. Overhead vapors from the stripper transfer energy to the rich solution before it enters the stripper.

The high pressure stripper shifts the equilibrium of the stripping reaction, thereby removing more ammonia from the rich solution. The resulting lean solution is recycled to the tray section of the superstill absorber. Steam at 600 psig provides the heat for the stripper reboiler.

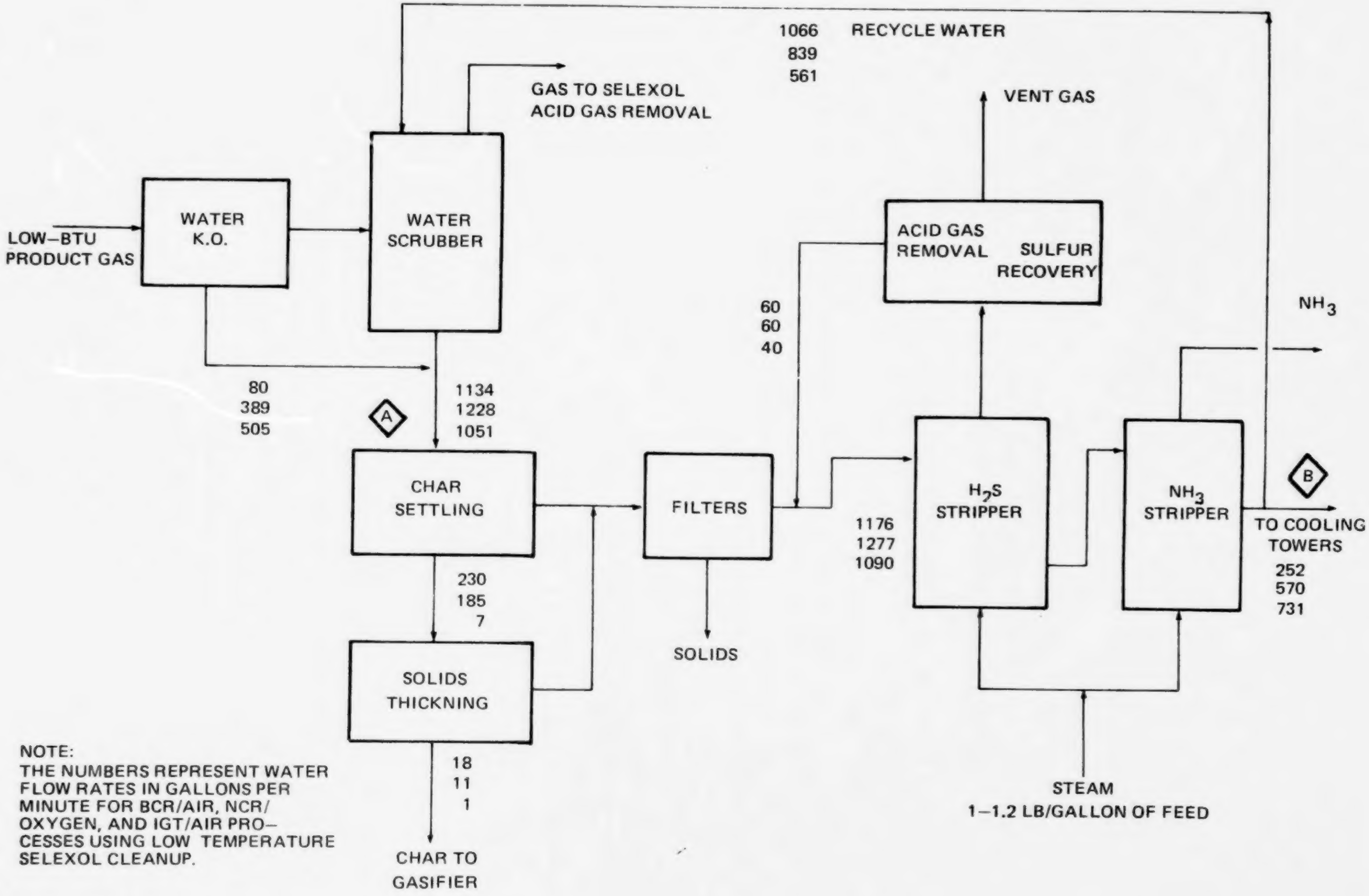
The vapor exiting the top of the stripper is 10 to 20 weight percent ammonia. The vapor is fed to a two-stage condenser. The first condenser stage heats the stripper feed. Cooling water absorbs heat energy in the second condenser stage. The resulting condensate passes to the fractionator feed tank from which it is pumped to the fractionator. The aqua-ammonia condensate is distilled into a 99.99 percent pure ammonia overhead fraction and a 0.05 percent ammonia bottoms fraction. Accumulation of trace quantities of acid gases and organics in the fractionator is prevented by adding caustic sodium hydroxide solution to the system, thereby assuring a high purity product. Hot water exiting the bottom of the fractionator is pumped to the bottom of the superstill to provide energy.

Sour Water Treatment Configuration and Cost

The generation, pretreatment and stripping of sour water for the combined cycle systems are shown in Figure A-5. Part of the unreacted feed steam condenses while passing through the heat recovery system and is separated from the gas in water knock-out drums. The gaseous mixture is then scrubbed with recycle water to remove ammonia. Ammonia removal is necessary to reduce the total plant NO_x producton during low-Btu fuel gas combustion. With the ammonia, some H_2S and CO_2 also dissolve in the scrubbing water. Entrained char particles will be removed by the scrubbing water. Removal of char particles is necessary to limit absorption on the turbo-generator blades and to reduce particulate emissions during low-Btu gas combustion. Water collected in the knockout drums and scrubber water is passed through gravity settling tanks to remove char particles. The solids-free stream is decanted, filtered, then passed to the stripping columns. Water condensate from the acid gas removal section also will be sent to the stripping columns. H_2S first will be stripped, then sent to the sulfur recovery process. Ammonia is stripped off in the separate column and can be recovered as aqueous ammonia by quenching it with water. About 1 to 2 pounds of steam will be required for H_2S and ammonia removal. The treated water will be free of suspended particles, ammonia and hydrogen sulfide. Part of the treated water is recycled to the scrubber and the rest is utilized in cooling towers.

FIG. A–5

SOUR WATER GENERATION AND TREATMENT


LOW–BTU PRODUCT GAS
WATER K.O.
WATER SCRUBBER
GAS TO SELEXOL ACID GAS REMOVAL
1066 RECYCLE WATER
839
561
VENT GAS
ACID GAS REMOVAL
SULFUR RECOVERY
60
60
40
NH3
80
389
505
A
1134
1228
1051
CHAR SETTLING
FILTERS
H2S STRIPPER
NH3 STRIPPER
B
TO COOLING TOWERS
252
570
731
1176
1277
1090
230
185
7
SOLIDS
SOLIDS THICKNING
18
11
1
CHAR TO GASIFIER
STEAM
1–1.2 LB/GALLON OF FEED

NOTE:
THE NUMBERS REPRESENT WATER FLOW RATES IN GALLONS PER MINUTE FOR BCR/AIR, NCR/OXYGEN, AND IGT/AIR PROCESSES USING LOW TEMPERATURE SELEXOL CLEANUP.

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The characterization of the inlet and outlet streams as shown in Figure A-5 is given in Tables A-7 to 9. The characteristics of the streams are based on the data published in the Office of Coal Research (OCR) report on coal gasification for combined cycle pilot plants. The cyanide, phenol, and organic concentrations of the sour water in this report were based on complex calculations of thermal equilibrium determining trace compounds in the gasifier produce gas. All of these components were assumed, as a worst case, to be absorbed into the plant sour water.

The capital cost for ammonia separation by the Two Stage Distillation Process and the Phosam-W Process for the three processes is shown in Table A-10. These costs are based on the cost figures estimated for different throughput by Water Purification Associates (Reference A-1).

BOILER FEEDWATER TREATMENT

Raw water initially will be pumped to a storage reservoir. The reservoir also serves as a flow equalizer and a clarifier which removes the suspended solids present in the natural water. At this stage the water going to the process is split into two streams, the smaller going to the demineralizers and the larger one to the cooling towers. The amount of demineralized water required depends on the total steam consumption of the process. Steam is required for gasification, sour water stripping and sulfur recovery. Boiler and demineralizer blowdowns have to be considered when estimating the boiler feed water stream. In general, the raw water to demineralizers can be estimated by summing up the total steam consumption and the boiler and demineralizer blowdowns.

The deminerlizers are used in the treatment scheme to remove ionic species and silica from boiler feedwater by exchanging these species with hydrogen, sodium and hydroxyl ions on solid resins. The resins are then regenerated by H_2SO_4, NaCl and NaOH, enabling their continued use and reuse over long periods. The most widely used ion exchange materials are resins of styrene-divinyl-benzene copolymers. The ion exchange resins play an important role in determining the process economics and demineralization capabilities. A summary of pertinent data for conventional ion exchange resins (Reference A-2) is presented in Table A-11. Also, a conventional demineralizer system is shown in Figure A-6.

There are a number of schemes that can be considered to optimize water demineralization. Two schemes are presented in Figure A-7. In Scheme 1, Ca^{++}, Mg^{++}, HCO_3^- and SiO_2 are removed by a hot lime treatment. Ca^{++} is further removed by a softening resin. Demineralization is carried out in successive steps with strong acid and weak base exchange followed by a mixed bed. In Scheme 2, the water is first treated with a weak acid ion exchange resin to remove Ca^{++}, Mg^{++} and some Na^+, and to replace these ions with H^+. The hydrogen ions release CO_2 from HCO_3^-, which is removed in the degasifier. Additional removal of Ca^{++} is required with a softening resin. Demineralization requires a strong acid resin followed by a weak base resin. A strong base resin is used for removal of SiO_2 and a mixed bed system is used for final polishing. Scheme 2 makes more efficient use of regeneration chemicals and costs less. A return condensate polishing system will be

TABLE A-7

FLOW CHARACTERIZATION FOR BCR/AIR-BLOWN/SELEXOL SOUR WATER TREATMENT
(SEE FIG. A-5)

	A		B	
STREAM COMPONENT	LB/HR	WT PPM	LB/HR	WT PPM
H_2O	235,720	--	125,950	--
H_2S	1,030	2,811	3	25
NH_3	8,154	22,260	6	50
CO_2	19,743	58,898	3	25
Phenols	3	8	1	< 8
Organics	12	33	4	< 30
Cyanides	7	19	1	< 5
Suspended Solids	2,880	7,862	1	< 5
Non-Condensibles	38	103	--	--
TOTAL	267,389		125,963	

TABLE A-8

FLOW CHARACTERIZATION FOR BCR/OXYGEN-BLOWN/SELEXOL SOUR WATER TREATMENT
(SEE FIG. A-5)

STREAM COMPONENT	A		B	
	LB/HR	WT PPM	LB/HR	WT PPM
H_2O	613,750	--	284,890	--
H_2S	1,080	1,670	6	21
NH_3	8,390	12,980	12	42
CO_2	20,716	32,044	6	21
Phenols	4	6	1.7	< 6
Organics	20	30	--	< 30
Cyanides	12	18	--	< 5
Suspended Solids	2,445	3,780	--	< 5
Non-Condensibles	64	100	--	--
TOTAL	646,481		284,916	

TABLE A-9

FLOW CHARACTERIZATION FOR IGT/AIR-BLOWN/SELEXOL SOUR WATER TREATMENT
(SEE FIG. A-5)

	A		B	
STREAM COMPONENT	LB/HR	WT PPM	LB/HR	WT PPM
H_2O	525,290	--	365,354	--
H_2S	337	632	1	3
NH_3	701	1,315	2	6
CO_2	6,284	11,793	1	3
Phenols	5	9	--	< 5
Organics	25	45	--	< 3
Cyanides	10	18	--	< 5
Suspended Solids	175	315	--	< 5
Non-Condensibles	35	63	--	--
TOTAL	532,867		365,358	

TABLE A-10

COST ESTIMATION OF SOUR WATER TREATMENT

COGAS Process	BCR/Air-Blown/Selexol	BCR/Oxygen-Blown/Selexol	IGT/Air-Blown/Selexol
Throughput, 10^6 gal/day	1.5	1.7	1.4
Two Stage Distillation Process			
Total Equipment Cost, 10^6 \$	3.9	4.2	3.5
Phosam-W Process			
Total Equipment Cost, 10^6 \$	3.1	3.1	2.6
Operating Costs for Phosam-W Process			
		Costs (\$/thousand gallons)	
Capital charges including maintenance at 17% yr (330 days/yr)	1.94	1.76	1.33
Energy at \$1.80/10^6 Btu	3.06	3.06	3.06
Caustic soda	0.04	0.04	0.04
Phosphoric acid	0.01	0.01	0.01
	5.05	4.87	4.44

TABLE A-11

SUMMARY OF PERTINENT DATA FOR CONVENTIONAL ION EXCHANGE RESINS

Ion Exchange Resins	Discussion	Flow Rate		Regeneration			Typical Ion Typical Capacity	
		[(m^3/day)/m^2]	(gpm/ft^3)	Chemical	(kg/m^3)	(lb/ft^3)	(Kg/m^3)	(Kgr/ft^3)
Cation Exchange Resins (Hydrogen or Acid Form)								
Sulfonic Acid (Strong Acid)	All feed water cations are removed	4-5	8-10	H_2SO_4	80	5	25	11
				H_2SO_4	176	11	57	25
				HCl	160	10	69	30
Carboxylic Acid (Weak Acid)	Cation removal is limited to quantities equivalent to the amount of weak anion (bicarbonate) that is present in the feed water.	3	6	H_2SO_4	160	10	137	60
				HCl	112	7	137	60
Anion Exchange Resins (Hydroxyl or Base Form)								
Strong Base	Anions of both strong and weak mineral acids are subject to greater physical degradation than are macro porous resins							
Type I - Gel Type		3	6	NaOH	64	4.0	25	11
Type II - Gel Type		3	6	NaOH	64	4.0	48	21
Type I - Marco Process		3	6	NaOH	64	4.0	23	10
Type II - Marco Process		3	6	NaOH	64	4.0	41	18
Weak Base Aliphatic Amine	Anions of strong mineral acids are removed. Anions of weak mineral acids such as CO_2, SiO_2 are not removed.	3	6	NaOH	59	3.7	66	29
				NH_4OH	51	3.2	66	29
				Na_2CO_3	79	4.9	66	29
Styrene Copolymer		4	8	NaOH	48	3.0	48	21
				NH_4OH	43	2.7	48	21
				Na_2CO_3	64	4.0	48	21

CONVENTIONAL TWO–BED ION EXCHANGE PROCESS

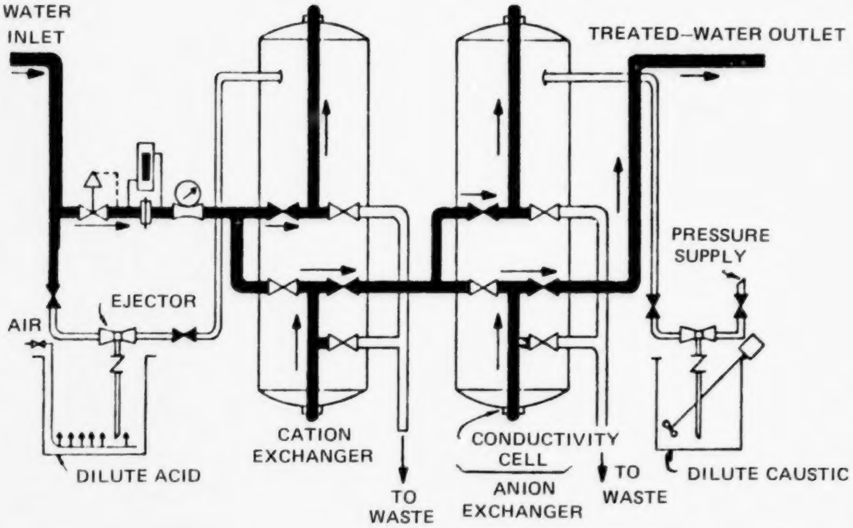
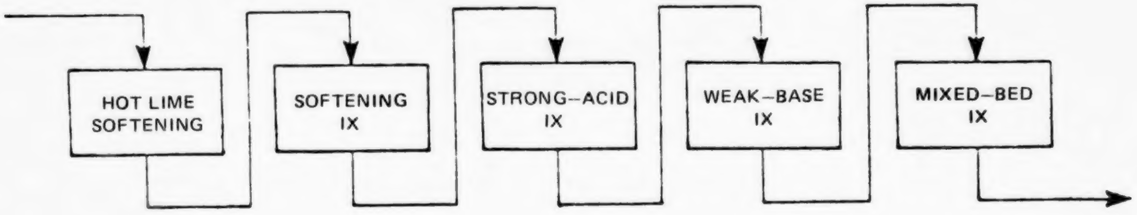
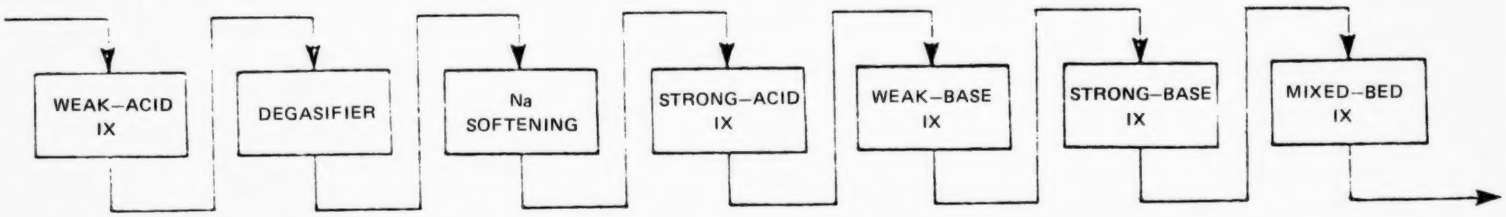

WATER INLET
TREATED–WATER OUTLET
EJECTOR
AIR
DILUTE ACID
CATION EXCHANGER
TO WASTE
CONDUCTIVITY CELL
ANION EXCHANGER
TO WASTE
PRESSURE SUPPLY
DILUTE CAUSTIC

FIG. A–7

BOILER FEED WATER TREATMENT SCHEMES


HOT LIME SOFTENING
SOFTENING IX
STRONG–ACID IX
WEAK–BASE IX
MIXED–BED IX

SCHEME I


WEAK–ACID IX
DEGASIFIER
Na SOFTENING
STRONG–ACID IX
WEAK–BASE IX
STRONG–BASE IX
MIXED–BED IX

SCHEME 2

IX = ION EXCHANGER

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required, but this is inexpensive. A raw water analysis is shown in Table A-12, the performance characterisitics of Scheme 2 are shown in Table A-13.

The deionized water will be used to generate steam at various pressures. The gasifier requires steam at 420 psig. Acid gas removal and ammonia recovery utilize steam at 50 psig and at 420 psig. The steam turbines will require steam at pressures above 1000 psig. The quality requirements for boiler feedwater at different pressures are shown in Table A-14. The deionized water produced by the Scheme 2 as mentioned above meets these boiler feedwater quality requirements.

Boiler Feedwater Treatment Cost

A cost summary for Scheme 2 demineralization, including the cost of chemicals for the three processes considered, is shown in Table A-15. The following costs are given for regenerating chemicals (¢ = cents): H_2SO_4, 3.8 ¢/lb; NaOH, 8.3 ¢/lb; and NaCl, 2.0 ¢/lb. The treatment costs were estimated by using the cost data published by Water Purification Associates (Reference A-1) (the treatment by Scheme 2 costs about \$3.2 x 10^5 for a flow rate of 869,000 lb/hr including resin replacement).

Reverse Osmosis for Boiler Feedwater Treatment

Reverse osmosis systems combined with demineralizers have proven to be cost effective in power plant feedwater use. Such a combination scheme for boiler feedwater treatment is used by some utility companies with an observed savings in the operating cost of more than one-third the conventional system. With the reverse osmosis unit operating upstream, the demineralizers produce five to ten times more deionized water between regenerations, and require less manpower to operate and maintain. In addition, the chemical requirements are lowered, thereby waste-disposal problems are minimized. Ion exchange resins operate with improved performance and extended life.

Figure A-8 shows the conventional demineralizer system and the combined reverse osmosis (RO)/demineralizer system, with a flow rate of 420,000 gpm; Table A-16 compares the cost (Reference A-3) of the two systems. The savings in chemical costs alone justifies application of the combined system.

A new approach to boiler feedwater treatment is the use of RO without demineralizers. Such a system has been operating for more than two years at a Texas petrochemical plant. The system, which feeds the boiler operating at 600 psi, produces deionized water from brackish well water containing 550 ppm dissolved solids. The 110,000 gpd system provides a 90 percent rejection at 85 percent recovery. The operating costs have ranged between 60 cents and 70 cents per thousand gallons.

COOLING TOWER WATER TREATMENT

The total make-up water to the cooling tower is primarily fresh water from the source, but also includes recycled water from the settling tank, boiler blowdown and the stripped water from the ammonia stripping section.

TABLE A-12

RAW WATER ANALYSIS

Component	Composition	as $CaCO_3$
pH	7.6	
Total Dissolved Solids	400 mg/l	
Bicarbonate (HCO_3^-)	180 mg/l	147.6
Sulfate (SO_4^{2-})	90 mg/l	93.6
Chloride (Cl^-)	170 mg/l	239.7
Nitrate (NO_3^-)	4.2 mg/l	--
Calcium (Ca)	52 mg/l	130.0
Magnesium (Mg)	14 mg/l	57.4
Sodium and Potassium (Na, K)	85 mg/l	147.1
Iron (Fe)	0.7 mg/l	1.3
Silica (SiO_2)	8.8 mg/l	--
Dissolved Oxygen (O_2)	9.8 mg/l	--
Ammonia (NH_3)	2.5 mg/l	--
Specific conductivity at 25°C	1.1×10^{-3} mho	--

TABLE A-13

PERFORMANCE CHARACTERISTICS FOR SCHEME 2 BOILER FEED WATER DEMINERALIZATION*

Component	After Weak Acid IX and Degasifier	After Softening	After Strong Acid Weak-Base Demineralization	Acid Strong-Base IX Silica Removal
CA + Mg	20	0	0	0
Na	69	99	44	4
Cl	170	170	0	0
HCO_3	0	0	0	0
SO_4	90	90	1	0
SiO_2	9	9	9	<1

*All Values in ppm

TABLE A-14

QUALITY REQUIREMENTS FOR BOILER FEEDWATER(1)

	Boiler Feedwater Quality of water prior to the addition of substances used for internal conditioning			
	Industrial			Electric Utilities
Characteristics	Low pressure 0 to 150 psig	Intermediate pressure 150 to 700 psig	High pressure 700 to 1,500	1,500 to 5,000 psig
Silica (SiO_2)	30	10	0.7	0.01
Aluminum (Al)	5	0.1	0.01	0.01
Iron (Fe)	1	0.3	0.05	0.01
Manganese (Mn)	0.3	0.1	0.01	(4)
Calcium (Ca)	(2)	(3)	(3)	(4)
Magnesium (Mg)	(2)	(3)	(3)	(4)
Ammonia (NH_4)	0.1	0.1	0.1	0.7
Bicarbonate (HCO_3)	170	120	48	(4)
Sulfate (SO_4)	(2)	(2)	(2)	(4)
Chloride (Cl)	(2)	(2)	(2)	(4)
Dissolved solids	700	500	200	0.5
Copper (Cu)	0.5	0.05	0.05	0.01
Zinc (Zn)	(3)	(3)	(3)	(4)
Hardness ($CaCO_3$)	20	(3)	(3)	(3)
Free mineral acidity ($CaCO_3$)	(3)	(3)	(3)	(3)
Alkalinity ($CaCO_3$)	140	100	40	(3)
pH, units	8.0-10.0	8.2-10.0	8.2-9.0	8.8-9.2
Color, units	(2)	(2)	(2)	(2)
Organics:				
Methylene blue active substances.	1	1	0.5	(3)
Carbon tetrachloride extract.	1	1	0.5	(3)
Chemical oxygen demand (O_2)	5	5	0.5	(3)
Dissolved oxygen (O_2)	2.5	0.007	0.007	0.007
Temperature, F	(2)	(2)	(2)	(2)
Suspended solids	10	5	(2)	(3)

(1) Unless otherwise indicated, units are mg/l and values that normally should not be exceeded. No one water will have all the maximum values shown.

(2) Accepted as received (if meeting total solids or other limiting values): has never been a problem at concentrations encountered.

(3) Zero, not detectable by test.

(4) Controlled by treatment for other constituents.

TABLE A-15

COST SUMMARY FOR DEMINERALIZATION OF BOILER FEED WATER

Process	Demineralizer Capacity (gpm)	Demineralizers Annual Operating (Cost $)	H_2SO_4	Chemicals Required lb/hr		Total Operating Cost $
				NaOH	NaCl	
BCR/Air-Blown Selexol	886.5	163,250	160	76	48	275,520
BCR/Oxygen-Blown Selexol	1073.4	197,440	194	91	58	325,440
IGT/Air-Blown Selexol	1533.2	282,240	277	131	83	466,240

CONVENTIONAL DEMINERALIZER AND RO/DEMINERALIZER SYSTEMS FOR BOILER FEEDWATER TREATMENT

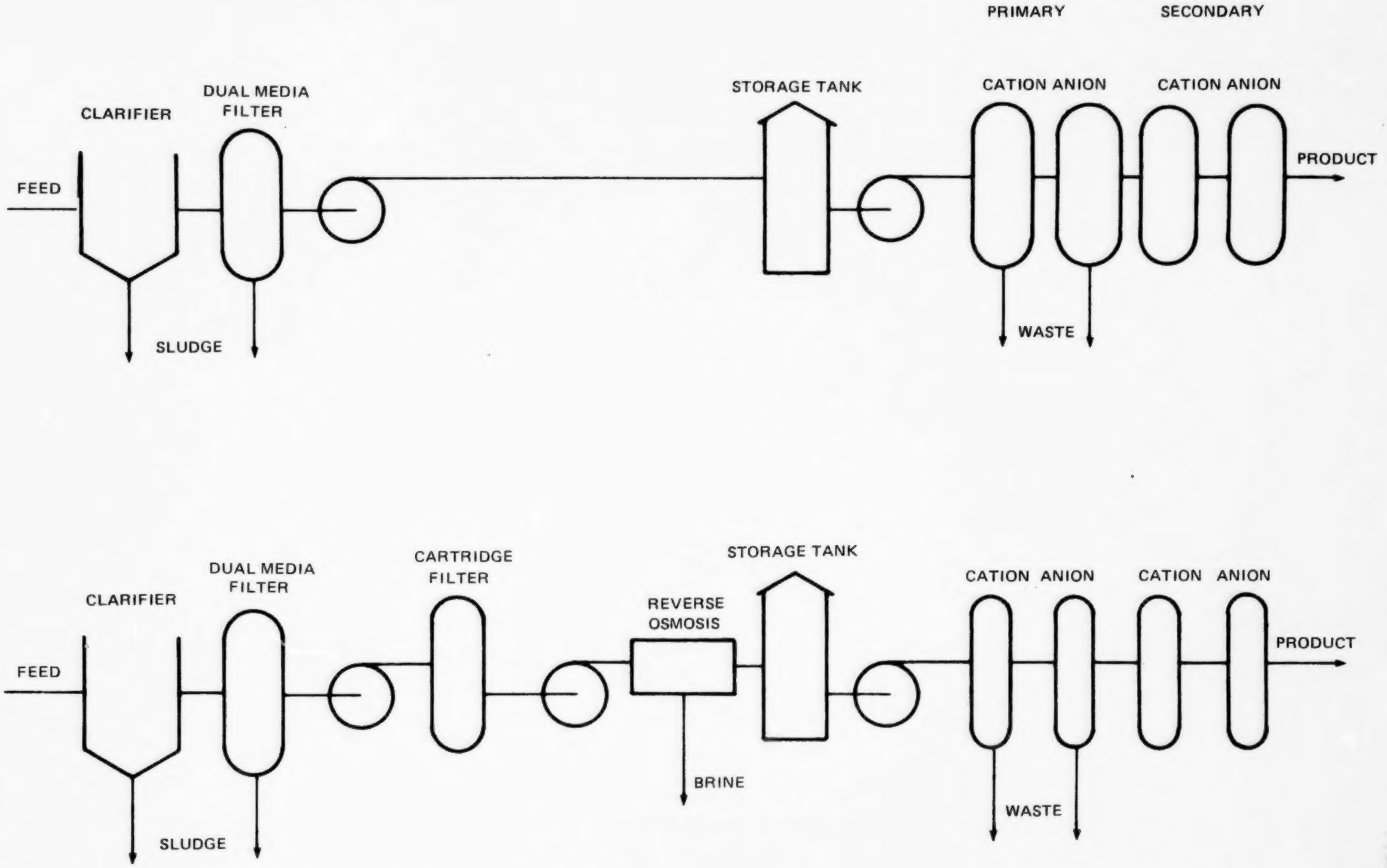
PRIMARY
SECONDARY
FEED
CLARIFIER
DUAL MEDIA FILTER
STORAGE TANK
CATION ANION
CATION ANION
PRODUCT
SLUDGE
WASTE
FEED
CLARIFIER
DUAL MEDIA FILTER
CARTRIDGE FILTER
REVERSE OSMOSIS
STORAGE TANK
CATION ANION
CATION ANION
PRODUCT
SLUDGE
BRINE
WASTE

FIG. A–8

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TABLE A-16

COST COMPARISON OF CONVENTIONAL DEMINERALIZER AND R/O DEMINERALIZER SYSTEMS

	Conventional		RO/Demineralizer	
	$/yr	$1000/gal	$/yr	$1000/gal
Capital	$160,230	1.015	$168,430	1.067
Labor	96,000	0.608	96,000	0.608
Maintenance	50,000	0.317	50,000	0.317
Pretreatment Chemicals	46,460	0.294	46,440	0.294
Demineralization, Chemicals	176,680	1.094	15,880	0.101
Resin	32,310	0.205	8,950	0.057
Modules	----	---	30,880	0.196
Waste Chemicals	81,300	0.515	2,570	0.016
Power	12,430	0.079	40,270	0.255
Total	651,410	4.13	459,420	2.91

Cost based on water containing 960 ppm TDS.

Cooling tower water is treated to prevent scaling, fouling, microbial growth and corrosion. The control limits for cooling tower circulating water are given in Table A-17.

The raw water may contain a high concentration of HCO_3^-, sometimes referred to as "alkalinity". Removal of bicarbonate prior to using the water in cooling towers will prevent scaling. The simplest way to accomplish this is to add sulfuric acid. One pound of HCO_3^- requires 0.8 lb H_2SO_4 for treatment. Although simple, this is an expensive treatment, in that it causes a waste of biocides and anticorrosion chemicals in the cooing tower blowdown.

Chemical precipitation is also a feasible pretreatment option for fresh water. This method reduces dissolved solids comprised of heavy-metal or hardness ions. Lime, soda-ash or caustic is added in softeners to precipitate the hardness. Cold lime softening is commonly employed. This restricts frequent cooling tower blowdowns and conserves the water conditioning chemicals.

The impact of pretreating cooling tower make-up is summarized in Table A-18. This table presents chemical analysis of lime-softened water, and recirculating water at 6 and 13 cycles of concentration. The quality of make-up water shown in this table is different from the quality of raw water considered. The analyses for both cases favor sulfuric acid addition for control of calcium carbonate. Calcium sulfate solubility, therefore, limits the cooling water concentration. Note that the scaling potential of calcium sulfate (as indicated by the solubility product of total calcium and sulfate in solution) is identical for each case by virtue of softening pretreatment. Operation at 13 rather than 6 cycles of concentration in this example reduces the tower blowdown by approximately 57 percent.

Cooling Tower Make-up Water Treatment Costs

The cost of cooling water by simple sulfuric acid treatment and by lime pretreatment for the processes considered is summarized in Table A-19. The following cost data has been used.

Sulfuric Acid Treatment

H_2SO_4 required for removal of 1 lb of HCO_3 = 0.8 lb.

H_2SO_4 cost = 3.8 cents/pound

Biocides cost = $\$1/10^3$ gallons of blowdown

Lime Treatment

Lime cost = 2.7 cents/pound

Biocides cost = $\$0.50/10^3$ gallons of blowdown

Anticorrosion chemicals = $\$0.50/10^3$ gallons of blowdown

TABLE A-17

CONTROL LIMITS FOR COOLING TOWER CIRCULATING WATER COMPOSITION

	Limits		
Parameters	Minimum	Maximum	Remarks
Langelier Saturation Index	+0.5	+1.5	Nonchromate programs
Ryznar Stability Index	+6.5	+7.5	Nonchromate programs
pH	6.0	8.0	
Calcium, mg/l as $CaCO_3$	20-50	300	Nonchromate program
		400	Chromate program
Total iron, mg/l		0.5	
Manganese, mg/l		0.5	
Copper, mg/l		0.08	
Aluminum, mg/l		1	
Suflide, mg/l		5	
Silica, mg/l		150	For pH 7.5
		100	For pH 7.5
$(Ca) \cdot (SO_4)$, product		500,000	Both calcium and sulfate expressed as mg/l $CaCO_3$
Total dissolved solids, mg/l		2,500	
Conductivity, micromhos/cm		4,000	
Suspended solids, mg/l		100-150	

TABLE A-18

IMPACT OF SOFTENING COOLING TOWER MAKE-UP WATER

Constituent	As	Plant Makeup	Following Primary Treatment	Circulating Waters Concentrations at 6	at 13
Calcium	$CaCO_3$	232	49	1392	640
Magnesium	$CaCO_3$	135	121	810	1580
Sodium	Na^+	117	175	702	2409
Bicarbonate	$CaCO_3$	129	0	35	35
Carbonate	$CaCO_3$	-	35	-	-
Sulfate	$SO_4^=$	345	355	2685	5821
Chloride	Cl^-	94	94	564	1221
Silicate	SiO_2	17	10	-	-
Total Hardness	$CaCO_3$	367	170	2202	2200
pH	-	8-8.3	8-9	6.5-7	6.5-7
Turbidity	-	5-20	5-10JTU	-	-
K_{sCaSO_4}*	-	-	-	$3.73x10^6$	$3.73x10^6$

---

*K_{sCaSO_4} = solubility product of calcium sulfate
= $(C_{Ca^{++}})$ $(C_{SO_4^=})$
where $(C_{Ca^{++}})$ = total calcium in solution, ppm as $CaCO_3$
(C_{SO_4}) = total sulfate in solution, ppm as $SO_4^=$

TABLE A-19

COOLING TOWER WATER TREATMENT COST

Operating Cost $/hr	BCR/Air-Blown/ Selexol	BCR/Oxygen-Blown/ Selexol	IGT/Air-Blown/ Selexol
H_2SO_4 Treatment	62.3	52.4	57.1
Lime Treatment	38.6	36.5	36.6

Lime Treatment (Cont'd)

Flocculation chemicals = \$0.30/1000 gallons of blowdown

Dispersant chemicals = \$0.15/1000 gallons of blowdown

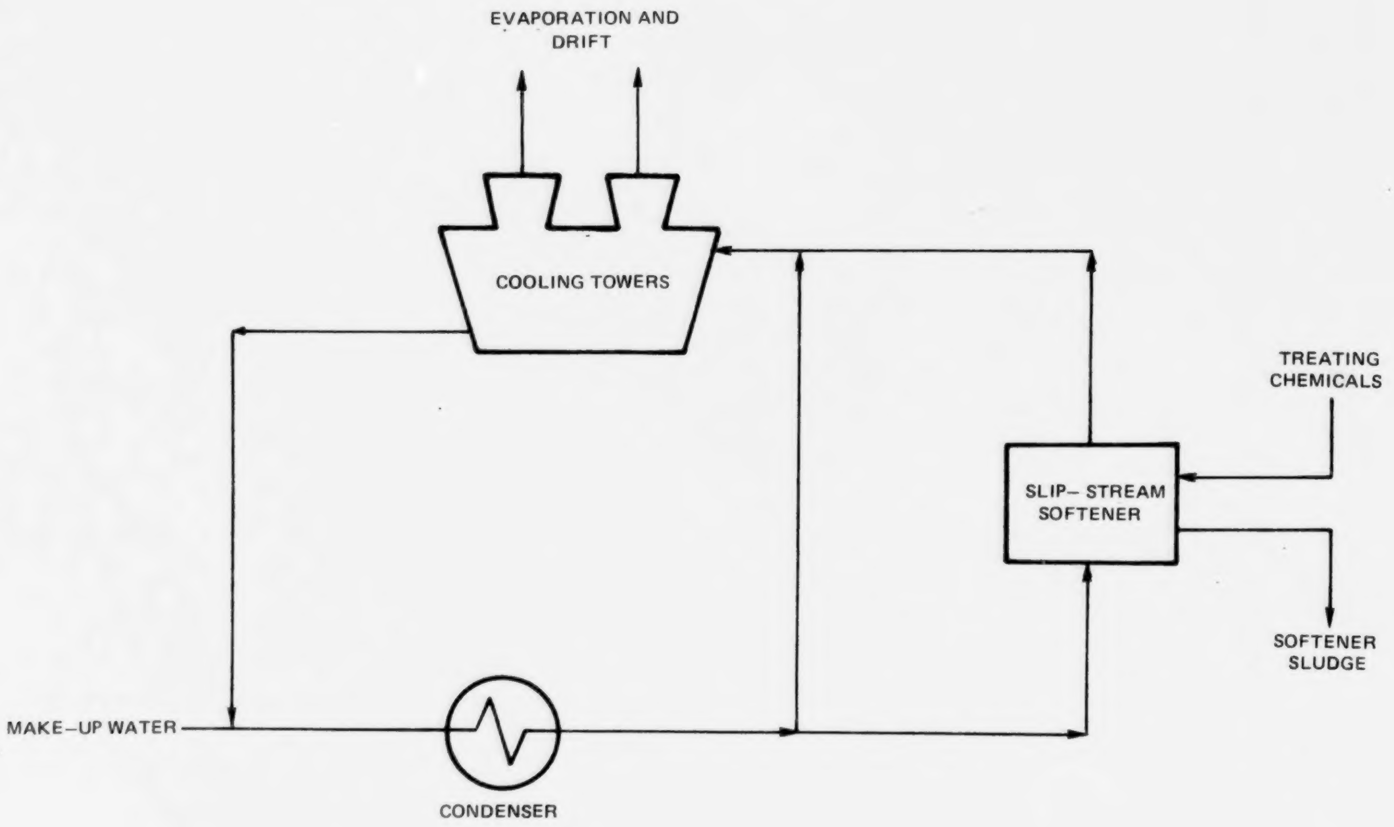
Table A-19 shows that the operating cost of cooling water treatment by lime pretreatment is about 60-70 percent of the cost of sulfuric acid treatment. However, lime treatment requires an initial capital investment of about one million dollars for the clarifiers and the side stream filter.

Sidestream Softening of Cooling Water

The amount of cooling tower blowdown can be greatly reduced by using side-stream treatment, such as lime softening and filteration. This method, shown in Figure A-9, treats the most concentrated water in the cooling tower. As a result, the amount of water treated is significantly less than the pretreatment of the make-up water. The system consists of an upflow, solids-contact reactor/clarifier. This is followed by filteration, and the treated water is recirculated to the cooling tower. Calcium and magnesium bicarbonates and silica concentration are reduced by lime treatment. Calcium and magnesium bicarbonates and silica concentration are reduced by lime treatment. Calcium and magnesium sulfates can be removed by adding soda ash together with lime. Treatment of the blowdown for removal of zinc, chromate, and other potential pollutants is much more practical when applied to a small flow. The performance of a typical system having a circulating water flow rate of 500,000 gpm and a side stream flow rate of 2000 gpm shows that the treatment reduces hardness to 80-120 mg/l, silica to 10-20 mg/l, and suspended solids to 10-20 mg/l. Lime treated effluent is recarbonated and then filtered to 1-2 mg/l of suspended solids. Sludge from the clarifiers is settled, dewatered on a filter and used for landfill.

FIG. A–9

SIDESTREAM SOFTENING OF COOLING WATER


EVAPORATION AND DRIFT
COOLING TOWERS
TREATING CHEMICALS
SLIP– STREAM SOFTENER
SOFTENER SLUDGE
MAKE–UP WATER
CONDENSER

78–02–138–14

REFERENCES

A-1 Goldstein, D. J. and David Yung: Water Conservation and Pollution Control in Coal Conversion Processes. EPA-600/7-77-065, (NTIS No. PB269-568), June, 1977.

A-2 Strauss, Sheldon D.: Water Treatment. Power, 117 (6) S1-S24 (1973).

A-3 Truby, Randy: Reverse Osmosis for Boiler Feedwater Treatment. Power Engineering, December, 1976, pp.58-60.

APPENDIX B

FUEL PROCESSING COST BASIS

Fuel processing costs for the low-temperature system were estimated by Fluor. The costs were estimated on a major component basis and an indication of the degree of detail considered is given in Table B-1. That table lists the major components that were considered in preparing an overall plant capital cost estimate.

TABLE B-1

EQUIPMENT LIST FOR AIR-BLOWN BCR PROCESS

Unit 10 - Coal Preparation and Handling

a. Coal Handling - one train required

Item	Description
10-BN-1	Unloading Hopper
10-CV-1	Belt Conveyor
10-CV-2	Belt Conveyor
10-TR-1	Tripper
10-ME-1	Stacking System
10-ME-2	Reclaiming system
10-CV-3	Belt Conveyor

b. Coal Pulverization - one train required

Item	Description
10-CV-4	Belt Conveyor
10-BN-2	Pulverizer Feed Hopper
10-1-V-1	Transport Gas
10-2-V-1	Heater
10-2-ME-3	Pulverization System
10-2-ME-3	
10-BN-3A, B, C & D	Pulverized Coal Storage Silos

Unit 20 - Gasification and Ash Handling

a. Gasification - five trains required (four operating and one spare) with each train consisting of the following:

Item	Description
Vessels	
20-1-V-1A & B	Coal Feed Lock Hoppers
20-1-V-2	High Pressure Coal Feed Hopper
20-1-V-3A & B	Slag Lock Hoppers
20-1-V-4	Char Separator
20-1-V-5	Char Cooler
20-1-V-6A & 6B	Char Transport Drums
20-1-V-7A & 7B	Slag Hoppers
Reactors	
20-1-R-1	Gasifier
Pumps	
20-1-P-1	HP Quench Water Pump
Compressors	
20-1-C-1	Transport Gas Compressor
Storage Bins	
20-1-BN-1	Coal Feed Surge Bin
Mechanical Equipment	
20-1-ME-1	Bag Filter

b. Ash Handling - one train required

Cooling Tower	
20-CT-1	Cooling Tower
Bins	
20-BN-2	Dewatering Bin

b. Ash handling - one train required (continued)

Item	Description
Tanks	
20-TK-1	Transfer Tank
20-TK-2	Settling Tank
Pumps	
20-P-2	Quench Water Supply Pump
20-P-3	Quench Water Return Pump
20-P-4	Settling Tank Bottoms Pump
20-P-5	Jet Ejector Water Pump
20-P-6	Transfer Tank Bottoms Pump

Unit 21 - Gas Cooling and Particulate Removal

a. Gas Cooling and Particulate Removal - high temperature gas cooling section consisting of 21-1-E-1, 21-1-E-2, 21-1-E-3, 21-1-V-1 and 21-1-P-1 have four operating and one spare trains and remainder of system has four operating trains and no spare. Each train consists of the following:

Vessels	
21-1-V-1	HP Steam Drum
21-1-V-2	Particulate Scrubber
21-1-V-3	K.O. Drum
21-1-V-4	Ammonia Absorber
Exchangers	
21-1-E-1	HP Boiler
21-1-E-2	Gasifier Effluent/Product Gas Exchanger
21-1-E-3	BFW Preheater
21-1-E-4	BFW Make-Up Heater
21-1-E-5	Demineralized Water Preheater
Pumps	
21-1-P-1	HP Boiler Circulation Pump
21-1-P-2	Particulate Scrubber Circulation Pump

b. Char Recovery - one train required

Item	Description
Pumps	
21-1-P-3	Char Slurry Recycle Pump
Mechanical Equipment	
21-ME-1	Secondary Char Recovery Unit

Unit 22 - Acid Gas Removal System

Data supplied by Allied Chemical Co.

Unit 23 - Sulfur Plant and Tail Gas Treating

a. Sulfur Plant

A total of three parallel trains (two operating and one spare) required with each train consisting of the following:

Furnace	
23-1-H-1	Sulfur Furnace and Waste Heat Boiler
Reactors	
23-1-R-1	Converter No.1
23-1-R-2	Converter No.2
Exchangers	
23-1-E-1	Condenser No. 1
23-1-E-2	Condenser No. 2
Blowers	
23-1-BL-1	Air Blower
Pumps	
23-1-P-1	Sulfur Transfer Pump
Sump	
23-1-S-1	Sulfur Sump

b. Tail Gas Treating

A total of three trains required (two operating and one spare) with each train consisting of the following:

Item	Description
Heater	
23-1-H-2	Tail Gas Heater
Reactor	
23-1-R-3	SO_2 Converter
Vessels	
23-1-V-1	Absorber/Quench Vessel
23-1-V-2	Sulfur Decanter
Exchangers	
23-1-E-3	Reactor Effluent Cooler
23-1-E-4	Quench Circulating Cooler
23-1-E-5	Regenerated Solution Cooler
Blowers	
23-1-BL-2	Air Blower
Pumps	
23-1-P-2	Quench Water Circulating Pump
23-1-P-3	Regenerated Solution Pump
23-1-P-4	Froth Slurry Pump
23-1-P-5	Filter Cake Pump
Tanks	
23-1-TK-1	Oxidizer Tank
23-1-TK-2	Froth Trank

Unit 24 - Sour Water Treating and Ammonia Recovery

One train required.

Item	Description
Vessels	
24-V-1	Superstill
24-V-2	Ammonia Stripper
24-V-3	Ammonia Fractionator
24-V-5	Fractionator Feed Tank
24-V-6	Flash Drum
Tanks	
24-TK-1	Phosphoric Acid Storage Tank
24-TK-2	Caustic Storage Tank
Exchangers	
24-E-1	Lean Solution Cooler
24-E-2	Solution Exchanger
24-E-3	Stripper Condenser
24-E-4	Condensate Subcooler
24-E-5	Fractionator Condenser
24-E-6	Superstill Reboiler
24-E-7	Absorber Cooler
24-E-8	Bottoms Cooler
24-E-9	Stripper Reboiler
24-E-10	Fractionator Reboiler
24-E-11	Reflux Condenser Cooler
24-E-12	Feed Preheater
Pumps	
24-P-1	Absorber Circulation Pump
24-P-2	Rich Solution Pump
24-P-3	Fractionator Feed Pump
24-P-4	Fractionator Reflux Pump
24-P-6	Superstill Bottoms Pump
24-P-7	Reflux Condenser Pump

APPENDIX C

POWERPLANT COST ANALYSIS

Previous studies demonstrated the need for a capability to provide cost estimates for future power systems that not only would enable power stations with different operating characteristics to be considered, but also would allow sensitivity studies to be undertaken for differing technical characteristics of the selected systems. As a result, members of the technical staff at UTRC, in cooperation with cost estimators at Burns & Roe, Inc., developed a set of correlations based on historical data which enable the total installed cost of COGAS power stations to be estimated in-house with a high degree of confidence. The procedure consists of a set of equations, grouped by Federal Power Commission (FPC) account numbers, which can be used to calculate the installed costs of major power station equipment.

Minor component equipment items are combined in logical groups, the costs of which are also estimated from correlation equations. The calculation procedure requires input data which are normally calculated as part of a routine technical analysis for electric power stations. Examples of such data include: gross and net station output power; steam turbine output power; gas turbine output power; cooling water flow rates; subcomponent input power requirements; and station efficiency. Because the cost correlations have been developed in sufficient detail, the need to resort to expensive preliminary system layout drawings is eliminated.

As part of the overall analysis, the costs of a few major components are estimated by methods other than the correlations described since these alternative methods make it possible to determine these specific equipment costs in even greater detail. Two such components are the large, industrial gas turbines and waste heat boilers, the costs of which are calculated using sophisticated computer program analyses proprietary to UTRC.

Furthermore, the costs of large, "standard" items such as steam turbogenerators and electric generators are obtained directly from published catalog data, which are corrected by appropriate price multiplier factors to maintain consistency with the latest industry experience.

A detailed list of all system components whose prices are capable of being estimated directly is presented in Table C-1 for all applicable FPC power station account categories. Not all items are used for each COGAS system considered since often there are specific characteristics of each station design which distinguish it from other stations. In the UTRC procedure, allowances are also made for such items as station start-up, temporary buildings, transportation during construction, special tools required during a particular power station contract, engineering, contingency, escalation (except where included in the cost of selected major system components), and interest during construction. Examples of typical resulting costs are shown to the right of each category. The example used is the air-blown BCR type gasifier with low-temperature cleanup which was the basic system of interest in this study.

TABLE C-1

POWER SYSTEM COST BREAKDOWN

FPC No. 341: Structures and Improvements

Site Preparation	1,209,000
Administration Building	805,000
Gas Turbine Building	1,088,000
Turbogenerator Building	3,738,000
Circulating Water System	1,438,000
Stack	1,724,000
	10,002,000

FPC No. 312 Boiler Plant Equipment

Waste Heat Boiler	56,365,000
Boiler Feed Pumps	741,000
Boiler Feedwater Tank and Deaerator	87,000
Demineralizer	906,000
Condensate Storage Tank	39,000
Misc. Pumps	156,000
Piping	12,265,000
Insulation for Piping	981,000
Controls	992,000
	72,532,000

FPC No. 314 Turbogenerator Unit

Turbogenerator	23,616,000
Condenser, Tube & Ejector	1,585,000
Condenser Vacuum Pumps & Motor	221,000
Condenser Pumps & Drive Motor	80,000
Cooling Tower	7,550,000
Circulating Water System	1,412,000
	34,464,000

TABLE C-1 (Cont'd)

Power System Cost Breakdown

FPC No. 343 Gas Turbine

Gas Turbine	46,038,000
Air Start System	180,000
Coupling to Generator	125,000
Lube Oil Purification System	137,000
Lube Oil Filtration	112,000
Turbine Air Precooler	603,000
Service Air Compressor	160,000
Breeching	3,540,000
Expansion Joints	2,064,000
Inlet Air Filters	472,000
Turbine Enclosure Air Cooler	120,000
Emergency Cooling Water System	12,000
Misc. Pumps & Tanks	28,000
Control Panels	300,000
Computer	300,000
Fuel Piping	1,501,000
Fuel Pipe Insulation	225,000
	55,917,000

FPC No. 341 Generator for Gas Turbine

14,863,000

FPC No. 345 Accessory Electrical Equipment
353 Station Equipment

17,620,000

FPC No. 346 Misc. Powerplant Equipment

526,000

NOT REPRODUCIBLE